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ENVIRONMENTAL PROTECTION AGENCY

REVIEW OF NEW SOURCE PERFORMANCE STANDARDS
FOR
PRIMARY COPPER SMELTERS

Prepared by:

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1. Existing standards of performance for primary copper smelters were promulgated in 1976. Section 111 of the Clean Air Act (42 USC 7411), as amended, directs that the Administrator periodically review promulgated standards.

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APPENDIX A

EVOLUTION OF THE REVIEW DOCUMENT

APPENDIX A

EVOLUTION OF THE REVIEW DOCUMENT

This study to review the existing standard of performance for primary copper smelters began in 1980, with Pacific Environmental Services, Inc. (PES). In September 1980, responsibility for the project was assigned to the Research Triangle Institute (RTI). Major events since RTI was assigned responsibility are shown in Table A-1.

Initial RTI activities include a review of the PES draft work plan and the preparation of the Phases II and III work plan. Discussions were held with PES and IERL/Cincinnati to identify and explicate the issues and to gather information documents for detailed study at RTI. In conjunction with EPA's Emission Monitoring Branch, a source test plan was prepared in June 1981. However, due to funding problems, source testing did not start until November 1981 with completion in January 1982. Radian Corporation performed the tests with RTI personnel observing.

Numerous plant visits were made during 1981 for familiarization and data collection purposes. Domestic smelters responded to 114 letters adding to the data base.

From September 1980 to date, numerous telephone and written contacts were made with foreign and domestic smelters, equipment suppliers, and domestic electric utilities to obtain information on primary copper smelter processes and emission control systems.

The technical background chapters describing the industry, emission control techniques, reconstruction and modification considerations, model plants, and regulatory alternatives were completed in March 1982, and mailed to industry for review and comment. The preliminary economic analysis was completed in September 1982 and the final economic analysis in October.

Industry comments on the draft BID were analyzed and incorporated into a revised version that was sent to working groups October 1982. Revised Chapters 6-9 were distributed to litigants and intervenors for review and comment in November 1982.

NAPCTAC review was accomplished in April 1983 and the notification package submitted for Steering Committee review and AA concurrence in September 1983.

TABLE A-1. MAJOR EVENTS AND ACCOMPLISHMENTS IN THE EVOLUTION OF THE BACKGROUND INFORMATION DOCUMENT

Month	Event
	Work begun by Pacific Environmental Services (PES). PES Work Plan submitted to EPA.
September 1980	Work begun by the Research Triangle Institute (RTI).
October 1980	Draft work plan discussed with I. J. Weisenberg, formerly project leader for PES effort.
October 1980	Draft Phases II and III Work Plan completed.
October 1980	Discussions with IERL/Cincinnati to identify issues and to obtain background documents.
December 1980	Phases II and III Work Plan completed.
February 1981	Familiarization visits made to five U.S. smelters--ASARCO/El Paso, Phelps Dodge/Hidalgo, Phelps Dodge/Morenci, Inspiration, and ASARCO/Hayden.
March 1981	Outokumpu Oy contacted and information obtained on the Outokumpu flash smelting system.
April 1981	Visits made to INCO Metals Company corporate headquarters and the Copper Cliff Smelter at Sudbury, Ontario, to assess capabilities of the INCO flash furnace.
May 1981	Familiarization visit made to Kennecott/Garfield smelter.
May 1981	Secondary air curtain for ASARCO/Tacoma converter discussed with ASARCO Engineering at Salt Lake City.
May 1981	Draft Source Test Plan completed.
June 1981	Source Test Plan completed.
July 1981	Pretest survey visits made to Phelps Dodge/Hidalgo and Phelps Dodge/Morenci.
September 1981	Visible emission tests conducted on converter secondary hoods at ASARCO/Tacoma.

(continued)

TABLE A-1 (continued)

Month	Event
November 1981	Tests conducted at Phelps Dodge/Hidalgo and Phelps Dodge/Morenci.
December 1981	Tests conducted on electric slag cleaning furnace scrubber and slag skim at Phelps Dodge/Hidalgo.
January 1982	Additional tests conducted on electric slag cleaning furnace at Phelps Dodge/Hidalga.
January 1982	Preliminary model plants defined.
March 1982	Technical background distributed for external review.
April 1982	Tabular cost data developed.
September 1982	Preliminary economic analyses completed.
October 1982	Cost study completed.
October 1982	Final economic analysis completed.
October 1982	Working group package distributed.
November 1982	Draft Chapters 6-9 distributed to litigants and intervenors for review and comment.
February 1983	NAPCTAC package distributed
April 1983	Review document reviewed by NAPCTAC
September 1983	Steering Committee package distributed
October 1983	Review document reviewed by NAPCTAC

APPENDIX B

INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

Table B-1 lists the locations in this document of certain information pertaining to environmental impact, as outlined in Agency Guidelines (39 FR 37419, October 21, 1974).

TABLE B-1. LOCATIONS OF INFORMATION CONCERNING ENVIRONMENTAL IMPACT WITHIN THE REVIEW DOCUMENT

Agency guidelines for preparing regulatory action environmental impact statements (39 FR 37419, October 21, 1974)	Location within the Review Document
Background and summary of emission control alternative	Chapter 6, Sections 6.2, 6.3, and 6.4
Statutory basis for review of the existing standard	Chapter 2, Section 2.1
Relationships to other regulatory agency actions	Chapters 3, 7, 9
Industry affected by the regulatory alternative	Chapter 3, Section 3.1, and Chapter 9, Section 9.1
Specific processes affected by the regulatory alternative	Chapter 3, Sections 3.2 and 3.6

APPENDIX B

INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

APPENDIX C
EMISSION SOURCE TEST DATA

APPENDIX C

EMISSION SOURCE TEST DATA

C.1 SUMMARY OF TEST DATA

EPA has undertaken several test programs in the past to assess the significance of and control techniques available for both process and fugitive SO_2 and particulate matter emissions from primary copper smelters. Portions of these data were used in this study and are summarized in Tables C-1 and C-2. For detailed discussions of these data, as well as discussions of the smelters involved in the previous testing programs, one may refer to either (1) the actual test reports from the U.S. Environmental Protection Agency's (EPA) Emission Measurement Branch (EMB), as presented in Tables C-1 and C-2, or (2) previously published EPA documents that have used the data--e.g., Arsenic Emissions from Primary Copper Smelters--Background Information for Proposed Standards, November 1980.

An additional test program was undertaken as a part of the current study to characterize smelter offgas streams for which data were scarce or nonexistent. Particulate matter and SO_2 mass emission rates were determined for several scenarios with combined EPA Reference Methods 5 and 6. Visible emissions data were also obtained for these sources with EPA Reference Method 9 and 22.

Brief discussions of each smelter and source tested during this study are presented in Sections C.2 and C.3, along with the test results.

A great deal of visible emissions data obtained during previous studies was used as reference material for this study. Therefore, for the reader's convenience, this data are presented in tabular form in Section C.4.

C.2 SUMMARY OF TESTING PERFORMED AT THE PHELPS DODGE-MORENCI SMELTER

At the time of testing, the Phelps Dodge-Morenci smelter had two reverberatory furnaces in operation, Nos. 3 and 5. Both furnaces were processing a green charge. The furnaces are fired with fuel oil.

Emissions tests were conducted to characterize matte tapping and slag skimming emissions from the Nos. 5 and 3 furnaces, respectively. Visual emissions data were also obtained to assess the effectiveness of the local hooding used to capture these emissions. The emissions test data are summarized in Table C-3, while the visual emissions data are summarized in Tables C-4 through C-6.

C.3 SUMMARY OF TESTING PERFORMED AT THE PHELPS DODGE-PLAYAS SMELTER

Several sources were tested at the Phelps Dodge-Playas smelter to characterize offgases associated with the operation of an Outokumpu flash smelter. Emissions tests were conducted to characterize offgases from flash furnace matte tapping and slag skimming, as well as offgases from electric slag cleaning furnace (ESGF) slag tapping. The primary offgas stream from the ESCF was also tested before and after particulate control by a wet venturi scrubber. These data are presented in Tables C-7 through C-9. Visual emissions data were also obtained for the tapping and skimming operations noted above. These data are presented in Tables C-10 and C-12.

C.4 SUMMARY OF VISIBLE EMISSIONS DATA OBTAINED PRIOR TO THE CURRENT REVISION

Many emissions data obtained by EPA were used in the current study. The data used are summarized in Table C-2 and detailed results are given in Tables C-13 through C-27.

TABLE C-1. SUMMARY OF EMISSION TEST RESULTS USED IN THE PRIMARY COPPER SMELTER NSPS REVIEW

Plant	Offgas source	Control equipment	Sampling location(s)	Sample type	Average particulate mass rate, kg/hr		Average SO_2 mass rate, kg/hr		EMB report no.
					Inlet	Outlet	Inlet	Outlet	
Anaconda[a] Anaconda, Montana	Fluid-bed roaster electric smelting furnace converter	Spray chamber/ baghouse	Inlet and outlet	Particulate matter	3,876	13.1	NA	NA	77-CUS-5
ASARCO[a] El Paso, Texas	Multihearth roasters and reverberatory furnace	Spray chamber/ cold ESP	Inlet and outlet	Particulate matter	1,129	37.2	NA	NA	78-CUS-7
	Reverberatory furnace matte tapping	Baghouse	Inlet	Particulate matter SO_2	1.2	NA	14.4	NA	78-CUS-7
	Converter building evacuation system	Baghouse	Inlet and outlet	Particulate matter SO_2	50.7	2.0	-	139	78-CUS-7
	Multihearth roaster discharge	-[b]	Primary offgas flue	Particulate matter SO_2	1.0	NA	2.4	NA	78-CUS-7
Phelps Dodge[a] Douglas, Arizona	Multihearth roaster discharge	Baghouse	Inlet and outlet	Particulate matter	285	1.2	NA	NA	78-CUS-8
Phelps Dodge[a] Ajo, Arizona	Reverberatory furnace matte tapping	-[b]	Fugitive gas flue	Particulate matter SO_2	2.2	NA	115	NA	78-CUS-9
	Converter blow cycle	-[b]	Fugitive gas flue	Particulate matter SO_2	27.7	NA	1,192	NA	78-CUS-9
Phelps Dodge Morenci, Arizona	Reverberatory furnace matte tapping	-[b]	Fugitive gas flue	Particulate matter SO_2	7.7	NA	136	NA	-
	Reverberatory furnace slag skimming	-[b]	Fugitive gas flue	Particulate matter SO_2	0.9	NA	7.7	NA	-
Phelps Dodge Playas, New Mexico	Flash furnace matte tapping[c]	-[b]	Fugitive gas flue	Particulate matter SO_2	20.4	NA	143.8	NA	81-CUS-8
	Flash furnace matte tapping[d]	-[b]	Fugitive gas flue	Particulate matter SO_2	2.9	NA	13.2	NA	81-CUS-8
	Flash furnace matte tapping[e]	-[b]	Fugitive gas flue	Particulate matter SO_2	3.9	NA	10.9	NA	81-CUS-8
	Flash furnace slag skimming	-[b]	Fugitive gas flue	Particulate matter SO_2	5.0	NA	59.0	NA	-
	Electric slag cleaning furnace	Particulate scrubber	Inlet and outlet	Particulate matter SO_2	45.4	2.2	81.6	44.9	-

[a] Test data obtained prior to this study.

[b] No control device used.

[c] At the flash furnace launder (without lancing emissions included).

[d] At the flash furnace doghouse enclosure (without lancing emissions included).

[e] At the flash furnace doghouse enclosure (with lancing emissions included).

TABLE C-2. SUMMARY OF VISIBLE EMISSIONS DATA USED IN THE PRIMARY COPPER SMELTER NSPS REVISION

Plant	Type of source	Methodology employed
ASARCO[a] Tacoma, Washington	Calcine discharge	EPA Method 22
	Matte tap port and launder	EPA Method 22
	Matte discharge into ladle	EPA Method 22
	Slag skim port and launder	EPA Method 9 EPA Method 22
	Slag discharge into pots	EPA Method 9 EPA Method 22
	Converter slag return	EPA Method 9 EPA Method 22
Phelps Dodge Morenci, Arizona	Matte tapping	EPA Method 9 EPA Method 22
	Slag skimming	EPA Method 9 EPA Method 22
Phelps Dodge Playas, New Mexico	Flash furnace matte tapping	EPA Method 9 EPA Method 22
	Slag skimming--electric slag cleaning furnace (ESCF)	EPA Method 9 EPA Method 22
	Matte tapping--electric slag cleaning furnace (ESCF)	EPA Method 9 EPA Method 22
	ESCF off-gas particulate scrubber	EPA Method 9
Tamano[a] Japan	Converter charging	EPA Method 9
	Converter copper blowing	EPA Method 9
	Converter slag blowing	EPA Method 9
	Converter slag pouring	EPA Method 9

[a] These data obtained prior to this study.

TABLE C-3. SUMMARY OF EMISSION RATES CALCULATED FROM PARTICULATE AND SULFUR DIOXIDE TESTING AT THE PHELPS DODGE-MORENCI SMELTER

	Estimated production		Particulate		Sulfur dioxide	
Source/test	Tons	Taps	lb/h[a]	lb/ton[b]	lb/h[a]	lb/ton[b]
Matte tapping (Reberb No. 5)						
EMB-004 MMT	185	8	19	0.1	290	1.6
EMB-006 MMT	250	10	18	0.072	290	1.2
EMB-008-MMT	275	11	15	0.054	310	1.1
Average			17	0.076	300	1.3
Slag skimming (Reverb. No. 3)						
EMB-003 MSS	80	2	2.0	0.025	15	0.19
EMB-005 MSS	90	3	2.5	0.038	30	0.33
EMB-007 MSS	60	2	1.2	0.020	7.6	0.13
Average			1.9	0.024	17	0.21

[a] lb of pollutant/h of sampling.

[b] lb of pollutant/ton of matte or slag produced during sampling.

TABLE C-4. VISIBLE EMISSION OBSERVATION DATA FOR REVERBERATORY FURNACE MATTE TAPPING OPERATIONS AT THE PHELPS DODGE-MORENCI SMELTER[a]

Duration of observation period, min	Average opacity for observation period, percent	Range of individual readings
8.75	8.57	5 to 25
8.50	2.06	0 to 25
6.50	8.85	5 to 20
8.50	8.09	5 to 30
5.00	7.25	5 to 10
6.50	7.31	5 to 20
9.00	11.39	5 to 20
11.00	15.68	5 to 30
9.50	16.71	10 to 20
4.00	10.00	5 to 10
9.50	14.20	5 to 30
6.50	18.46	10 to 30
9.50	47.06	10 to 60
8.00	17.34	10 to 40
5.00	6.88	5 to 25
7.75	18.23	10 to 30
5.00	17.75	10 to 30
7.50	14.50	5 to 35
5.00	7.00	0 to 30
9.25	24.86	10 to 70
6.50	7.50	0 to 30
3.75	6.67	0 to 30

[a] Based on visual observations made in accordance with EPA Method 9.

TABLE C-5. VISIBLE EMISSION DATA FOR REVERBERATORY FURNACE MATTE TAPPING OPERATIONS AT THE PHELPS DODGE-MORENCI SMELTER[a]

Duration of observation period, min	Percent of time emissions observed	Light reading, lux
6.0	100	350
7.0	100	175
5.0	82	350
5.0	100	88[b]

[a] Based on visual observations made in accordance with EPA Method 22.

[b] Not a valid observation since the light was less than 100 lux.

TABLE C-6. VISIBLE EMISSION OBSERVATION DATA FOR REVERBERATORY FURNACE SLAG SKIMMING OPERATIONS AT THE PHELPS DODGE-MORENCI SMELTER

Reference Method 9 results

Duration of observation period, min	Average opacity for observation period, min	Range of individual readings
30.00	0.00	-[a]
30.00	0.00	-[a]
33.00	2.72	0 to 5
6.25	11.00	5 to 30
27.00	0.00	-[b]
30.00	0.79	5 to 10

Reference Method 22 results

Duration of observation period, min	Percent of time emissions observed	Light reading, lux
30.00	3	175

---

[a] No opacity readings above 0.0 were observed.

TABLE C-7. SUMMARY OF EMISSION TEST RESULTS--MATTE TAPPING OF THE OUTOKUMPU FLASH FURNACE AT THE PHELPS DODGE-PLAYAS SMELTER

	Estimated production		Particulate		Sulfur dioxide	
Source/test	Tons	Taps	lb/h[a]	lb/ton[b]	lb/h[a]	lb/ton[b]
Matte tapping at the flash furnace launder[c]						
EMB-009 HMT	200	9	51	0.25	320	1.6
EMB-011 HMT	208	9	48	0.23	360	1.7
EMB-013 HMT	183	8	35	0.19	270	1.5
Average	197		45	0.22	317	1.6
Matte tapping at the flash furnace dog-house hooding[c]						
EMB-010 HDH	200	9	6.2	0.031	16	0.081
EMB-012 HDH	208	9	9.9	0.048	37	0.18
EMB-015 HDH	183	8	3.1	0.017	33	0.18
Average	197		6.4	0.032	29	0.15
Matte tapping at the flash furnace dog-house hooding[d]						
EMB-023 HDHL	144	7	8.6	0.060	24	0.16

[a] lb of pollutant/h of sampling.

[b] lb of pollutant/ton of matte tapped.

[c] Without lancing.

[d] With lancing.

TABLE C-8. SUMMARY OF EMISSIONS TESTS RESULTS--SLAG SKIMMING OF THE ELECTRIC SLAG CLEANING FURNACE AT THE PHELPS DODGE-PLAYAS SMELTER

Source/test	Estimated production		Particulate		Sulfur dioxide	
	Tons	Taps	lb/h[a]	lb/ton[b]	lb/h[a]	lb/ton[b]
Slag skimming						
EMB-054 HSS	142	3	11	0.075	120	0.86
EMB-055 HSS	140	4	10	0.073	150	1.10
EMB-056 HSS	180	6	12	0.069	120	0.68
Average	154		11	0.072	130	0.88

[a] lb of pollutant/h sampling.

[b] lb of pollutant/ton of slag skimmed.

TABLE C-9. SUMMARY OF EMISSION TEST RESULTS--ELECTRIC SLAG CLEANING FURNACE SCRUBBER AT THE PHELPS DODGE-PLAYAS SMELTER

Source/test	Sulfuric acid lb/h[a]	Particulate lb/h[a]	Sulfur dioxide lb/h[a]
Inlet			
EMB-016 HSI	-[b]	100	200
EMB-020 HSI	-[b]	110	150
EMB-022 HSI	-[b]	120	280
EMB-050 HSI8[c]	0.06	100	170
EMB-052 HSI8[c]	0.00	83	110
Average	0.03	100	180
Outlet			
EMB-017 HSO	-[b]	1.3	160
EMB-019 HSO	-[b]	0.98	63
EMB-021 HSO	-[b]	1.4	17
EMB-051 HSO8	0.10	19.0	70
EMB-053 HSO8	0.04	1.9	31
Average	0.07	4.9	99

[a] lb of pollutant/h of sampling.

[b] Results are to be considered only approximately representative of the scrubber conditions due to abnormal operation of the ESCF during the sampling period.

[c] Did not sample for sulfuric acid mist.

TABLE C-10. SUMMARY OF VISIBLE EMISSIONS DATA--MATTE TAPPING OF THE OUTOKUMPU FLASH FURNACE AT THE PHELPS DODGE-PLAYAS SMELTER

Number of taps observed[a,b]	Average opacity, percent	Total observation time, min:sec
3	20	32:00
8	20	55:00
1	20	21:00
6	40	30:00
1	30	4:00

Number of taps observed[c]	Percent of time emissions observed	Total observation time, min:sec
1	100	9:36
1	100	7:27
1	100	11:15
2	100	12:08

Number of taps observed[b,d]	Average opacity, percent	Total observation time, min:sec
1	30	13:00
1	35	10:00
1	45	9:00
2	40	23:00

Number of taps observed[c]	Percent of time emissions observed	Total observation time, min:sec
1	100	7:42
1	100	10:20
1	100	11:05

[a] Lancing emissions not included.

[b] Based on visual observations made in accordance with EPA Method 9.

[c] Based on visual observations made in accordance with EPA Method 22.

[d] Lancing emissions included.

TABLE C-11. SUMMARY OF VISIBLE EMISSIONS DATA--SLAG SKIMMING AND MATTE TAPPING OF THE ELECTRIC SLAG CLEANING FURNACE AT THE PHELPS DODGE-PLAYAS SMELTER

Operation	Summary
Slag skimming[a]	Method 9. Approximately 1.5 hours of opacity observations were made for two launders. The average opacity of fugitive emissions escaping one launder was 40 percent, while the average opacity of emissions from the other launder was less than 35 percent. Method 22. Two slag skimming launders were observed for a total of 108 minutes. Emissions escaped from one launder 99 percent of the time and from the other 81 percent of the time.
Matte tapping[a]	Method 9. Based upon 24 minutes of observation at a single launder, the fugitive emissions escaping capture had an average opacity of 45 percent. Method 22. One launder was observed for approximately 11 minutes. During this period, fugitive emissions were escaping 82 percent of the time.

[a] Lancing emissions included.

TABLE C-12. SUMMARY OF VISIBLE EMISSIONS DATA--ESCF OFFGAS PARTICULATE SCRUBBER

The scrubber is not a fugitive source; therefore, no Method 22 observations were performed.

Method 9. Based on a total of approximately 8.5 hours of observations, the average scrubber opacity was less than 5 percent.

TABLE C-13. VISIBLE EMISSION OBSERVATION DATA FOR ROASTER CALCINE DISCHARGE INTO LARRY CARS (EPA METHOD 22) AT ASARCO-TACOMA

		Observer 1		Observer 2			
Run no.	Date	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Mean duration of operation, min:sec	Mean % time emissions observed
1	6/24	1:20	0	1:15	0	1:18	0
2	6/24	2:40	0	2:40	0	2:40	0
3	6/24	1:20	0	1:20	0	1:20	0
4	6/25	1:23	0	1:23	0	1:23	0
5	6/25	1:58	0	1:52	0	1:55	0
6	6/25	1:42	0	1:42	0	1:42	0
7	6/25	1:12	0	1:13	0	1:13	0
8	6/25	1:20	0	1:20	0	1:20	0
9	6/26	2:50	0	2:49	0	2:50	0
10	6/26	1:48	0	1:48	0	1:48	0
11	6/26	2:30	0			2:30	0
12	6/26	1:42	0			1:42	0
13	6/26	3:04	0			3:04	0
				Average		1:54	0

TABLE C-14. VISIBLE EMISSION OBSERVATION DATA FOR MATTE TAP PORT AND MATTE LAUNDER (EPA METHOD 22) AT ASARCO-TACOMA

		Observer 1		Observer 2			
Run no.[a]	Date	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Mean duration of operation, min:sec	Mean % time emissions observed
1	6/24	6:24	0	6:36	1	6:30	0.5
2	6/24	6:00	0	6:00	0	6:00	0
3	6/24	4:51	0	4:55	3	4:53	1.5
4	6/24	6:05	0	6:10	0	6:08	0
5[b]	6/24						
6	6/25	2:58	0			2:58	0
7	6/25	5:22	0	5:22	0	5:22	0
8	6/25	5:36	0	5:36	0	5:36	0
9	6/25	5:08	0	5:10	0	5:09	0
10	6/25	6:02	0	5:33	0	5:48	0
11	6/25	5:12	0	5:13	0	5:13	0
12	6/25			6:37	0	6:37	0
13	6/25	4:50	0	4:53	0	4:52	0
14	6/25	5:23	0	5:22	0	5:23	0
15	6/25	5:17	0	5:18	0	5:18	0
16[b]	6/25						
17	6/25	5:13	0			5:13	0
18	6/25	5:58	0			5:58	0
				Average		5:26	0.13

[a] Method 22 data for corresponding runs at the matte discharge into the ladle are presented in Table C-15.

[b] Observations were made only at the matte discharge into ladle; see Table C-15.

TABLE C-15. VISIBLE EMISSION OBSERVATION DATA FOR MATTE DISCHARGE INTO LADLE (EPA METHOD 22) AT ASARCO-TACOMA

		Observer 1		Observer 2			
Run no.[a]	Date	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Mean duration of operation, min:sec	Mean % time emissions observed
1	6/24	6:30	0			6:30	0
2	6/24	5:49	0	5:40	0	5:45	0
3	6/24	4:53	0	5:01	0	4:57	0
4	6/24	6:12	0	6:10	0	6:11	0
5	6/24			6:31	0	6:31	0
6[b]	6/25						
7	6/25	5:09	0	5:02	0	5:06	0
8	6/25	5:21	0	5:28	0	5:25	0
9	6/25	5:02	0	5:03	0	5:03	0
10	6/25	4:29	0	4:32	0	4:31	0
11	6/25	5:12	0	5:13	0	5:13	0
12	6/25	6:16	0			6:16	0
13	6/25	4:43	0	4:45	0	4:44	0
14	6/25	5:13	0	5:15	0	5:14	0
15	6/25	5:15	0	5:09	0	5:12	0
16	6/25	5:41	0	5:50	0	5:46	0
17[b]	6/25						
18[b]	6/25						
					Average	5:30	0

[a] Method 22 data for corresponding runs at the matte tap and launder are presented in Table C-14.

[b] Observations were made only at the matte tape and launder; see Table C-14.

TABLE C-16. VISIBLE EMISSION OBSERVATION DATA FOR SLAG TAPPING AT SLAG TAP PORT AND SLAG LAUNDER (EPA METHOD 22) AT ASARCO-TACOMA

		Observer 1		Observer 2			
Run no.[a]	Date	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Mean duration of operation, min:sec	Mean % time emissions observed
1	6/24	12:25[b]	98[b]	12:26[b]	99[b]	12:26	
2	6/24	22:00	15	21:36	0	21:43	8
3	6/24	14:07	35[b]	13:52[b]	97[b]	14:07	
4[c]	6/24	14:10	13			14:10	13
5	6/25	16:44	11			16:44	11
6[c]	6/25	17:26	2			17:26	2
7	6/26	16:14	1			16:41	1
8	6/26	13:45	0.3			13:45	0
9	6/26	15:45	0			15:45	0
10	6/26	14:29	0			14:29	0
11[d]							
				Average		15:40	4
				Std. dev.			11

---

[a] Method 22 data for corresponding runs at the slag skim discharge point appear in Table C-18.

[b] Observations were made at the entire slag tap process line including the slag tap port, slag launder, and slag discharge into ladle, and therefore are not included in computing the mean of observations.

[c] Method 9 data for corresponding runs appear in Table C-17.

TABLE C-17. VISIBLE EMISSION OBSERVATION DATA FOR SLAG TAPPING AT SLAG TAP PORT AND SLAG LAUNDER (EPA METHOD 9) AT ASARCO-TACOMA[a]

Run no.	Date	Duration of operation, min.	Mean opacity, %	Mean opacity, %
1	6/25	14.75	1.3	10
2	6/25	18	10.3	30
Average		16.38	6	
Maximum				30

[a] Emission data were taken during entire slag tapping operation.

TABLE C-18. VISIBLE EMISSION OBSERVATION DATA FOR SLAG TAPPING--SLAG DISCHARGE INTO POTS (EPA METHOD 22) AT ASARCO-TACOMA

		Observer 1		Observer 2			
Run no.[a,b]	Date	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Mean duration of operation, min:sec	Mean % time emissions observed
1	6/24	12:46	97	12:26	73	12:36	85
2	6/24	21:09	93	21:43	99	21:26	96
3	6/24	14:06	97	13:52	95	13:59	96
4	6/24	14:05	82			14:05	82
5	6/25	16:34	91			16:34	91
6	6/25	17:29	94			17:29	94
7	6/26	15:54	90			15:54	90
8	6/26	13:48	86			13:48	86
9	6/26	15:48	77			15:48	77
10	6/26	14:11	72			14:11	72
11	6/26	14:45	82			14:45	82
				Average		15:31	86
				Std. dev.			8

[a] Visible emission observation data by EPA Method 9 for corresponding runs are presented in Table C-19.

[b] Visible emission observation data for corresponding runs for the slag tap port and launder are presented in Table C-16.

TABLE C-19. VISIBLE EMISSION OBSERVATION DATA FOR SLAG TAPPING AT SLAG DISCHARGE INTO POTS (EPA METHOD 9) AT ASARCO-TACOMA[a]

Run no.[b]	Date	Duration of operation, min	Mean opacity, %	Mean opacity, %
1	6/24	c		
2	6/24	c		
3	6/24	c		
4	6/25	13.75	22.7	50
5	6/25	16.75	11.3	30
6	6/25	11.75[d]	16	35
7	6/26	15	14.8	40
8	6/26	15	10.3	20
9	6/26	13	5.5	10
10	6/26	15	3.7	10
11	6/26			
Average		14.32	12	
Maximum				50

[a] Emission data were taken during entire slag tapping operation.

[b] Method 22 data for corresponding runs appear in Table C-18.

[c] No data were obtained by Method 9.

[d] Reading started after filling of first slag pot.

TABLE C-20. VISIBLE EMISSION OBSERVATION DATA FOR CONVERTER SLAG RETURN TO REVERBERATORY FURNACE (EPA METHOD 22) AT ASARCO-TACOMA

		Observer 1		Observer 2		Observer 3			
Run no.[a]	Date	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Duration of operation, min:sec	% time emissions observed	Mean duration of operation, min:sec	Mean % time emissions observed
1	6/24	1:04	100	1:05	89	0:58	100	1:04	96
2	6/24	0:47	97	0:47	96	0:46	100	0:46	98
3	6/24	0:54	100	0:53	100	0:55	100	0:53	100
4	6/25	0:55	100					0:55	100
5	6/25			1:03	100			1:03	100
6	6/25					0:52	100	0:52	100
7	6/25[b]								
8	6/26	1:04	66					1:04	66
9	6/26	1:00	85					1:00	85
10	6/26	1:15	83					1:15	83
11	6/26	0:55	82	0:41	93			0:48	88
12	6/26								
						Average		0:58	92
						Std. dev.			11

[a] Visible emission observation data by EPA Method 9 for corresponding runs are presented in Table C-21.

[b] No data obtained by Method 22.

TABLE C-21. VISIBLE EMISSION OBSERVATION DATA FOR CONVERTER SLAG RETURN TO REVERBERATORY FURNACE (EPA METHOD 9) AT ASARCO-TACOMA

		Observer 1			Observer 2		
Run no.	Date	Duration of operation, min:sec	Average opacity, %	Maximum opacity, %	Duration of operation, min:sec	Average opacity, %	Maximum opacity, %
1	6/24	[a]					
2	6/24	[a]					
3	6/24	[a]					
4	6/25	1.00	17.5	30	1.00	16	25
5	6/25	1.25	20	40			
6	6/25				1.00	23	35
7	6/25	0.75	23	35	0.75	23	30
8	6/26	1.25	5	10			
9	6/26	1.25	11	20			
10	6/26	1.50	12	20			
11	6/26	1.25	13	20			
12	6/26	0.75	5	10			

Average opacity for all readings--15%
Maximum opacity during all readings--40%

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[a] Data were not obtained by Method 9 on 6/24/80.

TABLE C-22. A SUMMARY OF METHOD 22 VISIBLE EMISSION OBSERVATION DATA FOR BLISTER DISCHARGE FROM CONVERTER AT THE TAMANO SMELTER IN JAPAN[a,b,c]

	Total time equal to or greater than given opacity							
	1st blister discharge		2nd blister discharge		3rd blister discharge		Total blister charge	
Opacity, %	min:sec	% of total time	min:sec	% of total time	min:sec	% of total time	min:sec	% of total time
5	8:00	53	11:30	96	1:00	29	20:30	67
10	5:00	33	8:45	73	0:30	14	14:15	47
15	3:15	22	5:15	44			8:30	28
20	1:30	10	3:15	27			4:45	16
25	0:30	3	2:00	17	0.15	7	2:45	8
30	0:15	2	0:45	6			1:00	3
35			0:15	2			0:15	<1

[a] Observation point: converter secondary hood system.

[b] Data were based on a total of 30.5-minute observations for three successive blister discharges of the total four blister discharges during one converter cycle. Duration of each of the three discharges observed was 15 minutes, 12 minutes, and 3.5 minutes, respectively.

[c] Table C-23 summarizes the observation data into average opacities for each set of 6-minute data.

[d] Total of the three individual blister discharges.

TABLE C-23. SUMMARY OF AVERAGE OPACITY FOR BLISTER POURING AT THE TAMANO SMELTER IN JAPAN[a]

Set no.[b]	Average opacity,[c] %
1	6
2	8
3	11
4	10
5	9

[a] Based on same observation data used for Table C-22.

[b] Observation time for each set is 6 minutes.

[c] Average of all sets is 9 percent.

TABLE C-24. SUMMARY OF METHOD 22 VISIBLE EMISSION DATA FOR INDIVIDUAL AND TOTAL MATTE CHARGES TO A CONVERTER OBSERVED AT THE TAMANO SMELTER IN JAPAN[a,b,c,d]

Opacity, %	Total time equal to or greater than given opacity									
	1st matte discharge		2nd matte discharge		3rd matte discharge		4th matte discharge		Total matte charge	
	min:sec	% of total time	min:sec	% of total time	min:sec	% of total time	min:sec	% of total time	min:sec	% of total time
5	0:45	43	0:45	60	0:45	43	0		2:15	35
10			0:15	20	0:15	14			0:30	8
25	0:15	14							0:15	4

[a] Matte Charges 1, 2, and 3 were successive charges; respective charging times for Matte Charges 1, 2, 3 and 4 were 1.75 min., 1.25 min., 1.75 min., and 1.75 min., respectively.

[b] Observation point: converter secondary hood system.

[c] Data are based on a total of 6.5-minute observations for three successive matte charges at the beginning of one converter cycle and an intermediate matte charging during the cycle. Average duration of each matte charge was 1.5 minutes.

[d] Total of the four individual matte charges; average opacity for matte charging, based on total observation, is 3.0 percent.

TABLE C-25. SUMMARY OF VISIBLE EMISSION OBSERVATION DATA FOR COPPER BLOW AT THE TAMANO SMELTER IN JAPAN[a]

Set no.[b]	Average opacity, %
1	0
2	0
3	0
4	0

[a] Observation point: converter secondary hood system.

[b] Each set is based on 6-minute observation.

TABLE C-26. SUMMARY OF VISIBLE EMISSION OBSERVATION DATA FOR SLAG BLOW AT THE TAMANO SMELTER IN JAPAN

Set no.[b]	Average opacity, %
1	0
2	0
3	0
4	0
5	0

[a] Observation point: converter secondary hood system.

[b] Each set is made up of 6-minute observation; first two sets of data are based on observations during 1st slag blow and the remaining three sets of data are based on observations during 2nd slag blow of the total three slag blows in a converter cycle at the Tamano smelter.

TABLE C-27. SUMMARY OF VISIBLE EMISSION OBSERVATION DATA FOR CONVERTER SLAG POURING AT THE TAMANO SMELTER IN JAPAN[a]

Set no.[b]	Average opacity, %
1	0
2	0

[a] Observation point: converter secondary hood system.

[b] Each of two consecutive sets of 6-minute observations are made during one slag discharge.

APPENDIX D

(NOT USED)

APPENDIX E

USE OF COAL IN THE OUTOKUMPU FLASH FURNACE AT THE TOYO SMELTER

APPENDIX E

USE OF COAL IN THE OUTOKUMPU FLASH FURNACE AT THE TOYO SMELTER

At the Toyo smelter in Japan,[1] additional heat is supplied to the flash furnace by preheated air, coal, and oil. This smelter is in the process of converting from oil to coal because of the lower price of the latter. The use of coal at the Toyo smelter began in April 1981 and has continued for over 9 months. Initially, pulverized coal was substituted for half of the oil requirement of the furnace. The coal is fed to each of the concentrate burners. The rate of coal addition is controlled carefully in order to control the matte grade of the furnace--the coal being combusted preferentially to the concentrate feed. Personnel at the Toyo smelter have reported that no problems have been encountered related to operation of the flash furnace, waste heat boiler, electrostatic precipitator, or acid plant since coal has been used.[1] Because of the successful operations, the conversion from oil to coal has proceeded at a greater rate than expected.

REFERENCE

1. Moriyama, K., T. Terayama, T. Hayashi, and T. Kimura. The Application of Pulverized Coal to the Flash Furnace at Toyo Smelter. In: Copper Smelting--An Update, George, D. B. and J. C. Taylor (eds.). Warrendale, PA, The Metallurgical Society of AIME. 1981. p. 201-212.

APPENDIX F

COST ANALYSIS TO ESTIMATE THE INCREMENTAL INCREASE IN CAPITAL COST INCURRED BY INCREASING SULFURIC ACID PLANT GAS-TO-GAS HEAT EXCHANGER CAPACITY

Case I: Incorporate additional heat exchanger capacity in the plant design to lower the autothermal operating requirement for a double contact/double absorption (DC/DA) plant from 4.0-percent SO_2 to 3.5-percent SO_2.

Based on an overall heat transfer coefficient, U, of 4.0 $\text{Btu/hr} \cdot \text{ft}^2 \cdot {}^\circ\text{F}$;*

Heat exchanger surface area required with a 4.0-percent SO_2 gas stream entering the acid plant converter $\cong$ 4.15 ft^2/scfm.

Heat exchanger surface area required with a 3.5-percent SO_2 gas stream entering the acid plant converter $\cong$ 5.70 ft^2/scfm.

Heat exchanger cost (mid-1980 dollars) = \$25.22/ft^2.

Indexing to mid-1981 dollars,† we have

$$\text{Heat exchanger cost} \cong \frac{\$25.22}{\text{ft}^2} \times \frac{717.9}{652} = \$27.77/\text{ft}^2 .$$

Thus, at 4-percent SO_2, the total heat exchanger cost for a DC/DA plant is estimated as:

$$\frac{4.15\ \text{ft}^2}{\text{scfm}} \times \frac{\$27.77}{\text{ft}^2} = \frac{\$115.25}{\text{scfm}} .$$

Similarly, at 3.5-percent SO_2, the total heat exchanger cost can be estimated to be \$158.29 per scfm.

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*Weisenberg, I. J., and T. Archer. "Feasibility of Primary Copper Smelter Weak SO_2 Stream Control Relative to Reverberatory Furnace NSPS Exemption," Draft Final Report, July 1978.

†Marshall and Swift Equipment Cost Indices, Chemical Engineering, February 8, 1982.

Thus, the incremental cost, Δ\$, is estimated as:

$$\Delta\$ = \$158.29 - \$115.25 = \$43.04 \text{ per scfm.}$$

The total installed capital cost for a DC/DA plant designed to operate autothermally at 4.0-percent SO_2 is presented in Figure 8-1. At 50,000 scfm, this cost is estimated to be \$26.21 MM. The increase in the installed capital cost (due to the increased heat exchanger capacity) required to lower the autothermal operating requirement to 3.5-percent SO_2 is estimated as follows:

$$\frac{\$43.04}{\text{scfm}} \times 50{,}000 \text{ scfm} = \$2{,}152{,}000.00 \ .$$

Thus, the increase in the installed capital cost incurred as a result of lowering the autothermal operating requirement from 4.0- to 3.5-percent SO_2 is calculated as follows:

$$\$28{,}362{,}000.00 = (1 + f) \times \$26{,}210{,}000.00 \ ,$$

where f = the fractional increase in the installed capital cost.

Solving for f yields

$$f = 0.082 \ .$$

Thus, as a result of lowering the autothermal requirement from 4.0- to 3.5-percent SO_2, the installed capital cost of the plant increases about 8.2 percent at the 50,000 scfm level.

Similarly, at the 200,000 scfm level, the installed capital cost would be expected to increase about 12.8 percent. Thus, over the 50,000 to 200,000 scfm range, reducing the autothermal operating requirement for a DC/DA plant from 4.0- to 3.5-percent SO_2 would be expected to increase the installed capital cost by 8.2 to 12.8 percent.

Case II: Incorporate additional heat exchanger capacity in the plant design to lower the autothermal operating requirement for an single contact/single absorption (SC/SA) plant from 3.5-percent SO_2 to 3.0-percent SO_2.

Based on an overall heat transfer coefficient, U, of 4.0 Btu/hr $\cdot$ ft^2 $\cdot$ °F,

Heat exchanger surface area required with a 3.5-percent SO_2 gas stream entering the acid plant converter ≅ 1.80 ft^2/scfm.

Heat exchanger surface area required with a 3.0-percent SO_2 gas stream entering the acid plant converter ≅ 2.45 ft^2/scfm.

Heat exchanger cost (mid-1980 dollars) = \$25.22/ft^2.

Indexing up to mid-1981 dollars yields a heat exchanger cost of \$27.77 per square foot.

Thus, at 3.5-percent SO_2, the total heat exchanger cost for an SC/SA plant is estimated as follows:

$$\frac{1.8\ \text{ft}^2}{\text{scfm}} \times \frac{\$27.77}{\text{ft}^2} = \frac{\$50.00}{\text{scfm}} .$$

Similarly, at 3.0-percent SO_2, the total heat exchanger cost can be estimated to be \$68.00 per scfm.

Thus, the incremental cost, Δ\$, is estimated as follows:

$$\Delta\$ = \$68.00 - \$50.00 = \$18.00 \text{ per scfm} .$$

The total installed capital cost for an SC/SA plant designed to operate autothermally at 3.5-percent SO_2 is presented in Figure 8-5. At 50,000 scfm, this cost is estimated to be \$22.68 MM. The increase in the installed capital cost (due to the increased heat exchanger capacity) required to lower the autothermal operating requirement to 3.0-percent SO_2 is estimated as follows:

$$\frac{\$18.00}{\text{scfm}} \times 50{,}000 \text{ scfm} = \$900{,}000 .$$

Thus, the increase in the installed capital cost incurred as a result of lowering the autothermal operating requirement from 3.5- to 3.0-percent SO_2 is calculated as:

$$\$23{,}580{,}000.00 = (1 + f) \times \$22{,}680{,}000.00 ,$$

where f = the fractional increase in the installed capital cost. Solving for f yields

$$f = 0.0397.$$

Thus, as a result of lowering the autothermal requirement from 3.5- to 3.0-percent SO_2, the installed capital cost of the plant increases about 4.0 percent at the 50,000 scfm level. Similarly, at the 200,000 scfm level the installed capital cost would be expected to increase about 6.4 percent. This, over the 50,000 to 200,000 scfm range, reducing the autothermal operating requirement for an SC/SA plant from 3.5- to 3.0-percent SO_2 would be expected to increase the installed capital cost by 4.0 to 6.4 percent.

APPENDIX F

COST ANALYSIS TO ESTIMATE THE INCREMENTAL INCREASE IN CAPITAL COST INCURRED BY INCREASING SULFURIC ACID PLANT GAS-TO-GAS HEAT EXCHANGER CAPACITY

APPENDIX G

ANALYSIS OF CONTINUOUS SO_2 MONITOR DATA AND DETERMINATION OF AN UPPER LIMIT FOR THE INCREASE IN SO_2 EMISSIONS DUE TO SULFURIC ACID PLANT CATALYST DETERIORATION

Please note: To provide the most comprehensive study possible, this appendix is reprinted, with minor editorial changes, from Volume I, Proposal Standards, of Background Information for New Source Performance Standards: Primary Copper, Lead, and Zinc Smelters, publication number EPA 450/2-74-002a.

APPENDIX G

ANALYSIS OF CONTINUOUS SO_2 MONITOR DATA AND DETERMINATION OF AN UPPER LIMIT FOR THE INCREASE IN SO_2 EMISSIONS DUE TO SULFURIC ACID PLANT CATALYST DETERIORATION

G.1 EMISSION VARIATION

SO_2 emissions from the No. 7 sulfuric acid plant, which is the newest of five single-stage absorption plants operating on the offgases from the nine Kennecott copper converters at Garfield, Utah, were analyzed. The emissions were recorded by a Du Pont 460 continuous SO_2 analyzer from September 15, 1972, to November 15, 1972. This instrument is capable of measuring SO_2 concentrations within ±150 ppm (2 percent of full scale) and automatically zeroes itself every 8½ minutes. The zero calibration procedure requires 1½ minutes; thus the instrument is "on-line" 85 percent of the time.

A general review of the data generated revealed that several periods of data were missing due to problems with the recorder. Other segments contained long periods of plant shutdowns for maintenance or included concentrations that were obviously greater than the upper limit of the monitor. (A shorter absorption tube could have been installed to increase the upper limit of the monitor, if this situation had been noticed sooner.) Consequently, on the basis of data legibility and continuity, the periods of October 11-27, 1972, and November 8-15, 1972, were selected as representative of the 2-month monitoring period.

Periods of emissions during which the average concentration appeared to be greater than 3,000 ppm or less than 1,000 ppm were then noted. Eighteen periods during which emissions exceeded 3,000 ppm, including two periods during which emissions exceeded the recording capacity of the Du Pont analyzer (7,500 ppm), were identified. Fourteen periods during which emissions were less than 1,000 ppm were also identified. Acid plant operating logs and inlet SO_2 volume and concentration continuous monitor data were analyzed to ascertain if upsets, malfunctions, or startups and shutdowns occurred during these periods.

One major upset/malfunction was discerned. It occurred during one of the two periods during which the emissions exceeded the recording capacity of the analyzer. The upset/malfunction resulted from prolonged low inlet SO_2 concentrations, which caused a decrease in the normal temperature increase across the first catalyst bed. Consequently, this period of excessive emissions was deleted from the data. Six shutdowns and startups were noted. The six periods of low emissions following these shutdowns were deleted from the data because the acid plant was not in operation. Two periods of high emissions were identified following two of the six startups. These two periods of high emissions were also deleted from the data. Due to the time constraints placed on the analysis of these data, no investigation of why four of these six startups had no associated periods of high emissions was conducted. A brief investigation of the eight remaining periods during which emissions were less than 1,000 ppm, however, did reveal that these low emissions appeared to be the result of almost ideal operating conditions within the acid plant, with somewhat low inlet gas volumes and SO_2 concentrations and a minimum of fluctuations in either of these variables.

Following this review of acid plant operating data, fifteen periods during which emissions were higher than 3,000 ppm remained. This included one of the two periods previously identified as periods during which emissions exceeded the capacity of the Du Pont analyzer. This period was then deleted from the data for the following reasons. First, and most important, because no knowledge concerning numerical values of emissions was available, this time period could not be mathematically accounted for in the analysis. Second, because emissions were apparently so great, this period of operation would represent a violation of any reasonable standard developed and thus would add nothing to the analysis of "normal" operating emissions data to provide a basis for such standards.

The long-term SO_2 emissions concentration average was then calculated for all the data generated during the "normal operating" portions of the October 11-27 and November 8-15 periods. Fifteen-minute instantaneous SO_2 concentration values were used for this calculation, and the long-term emission average was determined to be 1,700 ppm. It is significant to note that this value is considerably less than the emission concentration corresponding to Monsanto's guaranteed conversion efficiency of 95 percent conversion of SO_2 to SO_3 at 5 percent SO_2 inlet, i.e., approximately 2,700 ppm.

The 14 periods of high emissions that were not deleted from the data were then examined by averaging these periods over various time intervals using the 15-minute instantaneous SO_2 concentration values identified during the above analysis. The time-averaged concentrations were then compared to various outlet SO_2 concentration levels to determine the extent to which such averaging periods mask variations in outlet concentration. The results are tabulated in Tables G-1 and G-2.

Seven of the fourteen high-emission periods exceeded 2,700 ppm (equivalent to the manufacturer's guarantee) when averaged for a 6-hour duration. Increasing the averaging time to 7 hours decreased the number of periods exceeding 2,700 ppm to five. Further increases in the averaging period resulted in only minor decreases in the number of periods exceeding 2,700 ppm. Increasing the level of average SO_2 emission concentration from 2,700 ppm to 3,000 ppm (approximately 10 percent) caused a significant reduction of the number of high-emission periods that exceeded this level as compared with 2,700 ppm. For each time-averaging interval, the number of periods for which the averages exceed 3,000 ppm is about half the number of periods corresponding to 2,700 ppm. Increasing the level of average SO_2 emission concentration from 2,700 to 3,250 ppm (approximately 20 percent) resulted in only a slight decrease in the number of periods exceeding this level compared to the number of periods exceeding 3,000 ppm. In general, therefore, increasing either the averaging time to periods greater than 6 hours, or increasing the average SO_2 emission concentra-

TABLE G-1. SUMMARY OF PERIODS EXCEEDING THE REFERENCE LEVEL SO_2 CONCENTRATION AS A FUNCTION OF AVERAGING TIME

Concentration (ppm)	4-h average	6-h average	7-h average	8-h average	12-h average
2,700	13	7	5	5	3
3,000	8	4	3	3	1
3,250	5	3	3	2	0

TABLE G-2. SUMMARY OF TOTAL TIME EXCEEDING THE REFERENCE LEVEL SO_2 CONCENTRATION AS A FUNCTION OF AVERAGING TIME

Concentration (ppm)	4-h average	6-h average	7-h average	8-h average	12-h average
2,700	112 (21)	76 (14)	62 (11)	62 (11)	42 (8)
3,000	61 (11)	40 (7)	33 (6)	33 (6)	13 (2)
3,250	40 (7)	30 (6)	30 (6)	22 (4)	0 (0)

NOTE: Numbers in parentheses indicate percentage of time for which the emissions would exceed the reference concentration. The total "normal" operating time of 542 hours equals 100 percent.

tion selected for comparison by more than 10 percent above the manufacturer's guarantee, does not significantly decrease the number of high-emission periods that exceed the level of SO_2 emission concentration selected for comparison.

Another approach is to examine the actual time during which SO_2 emissions exceeded various selected concentration levels, such as 2,700, 3,000, and 3,250 ppm. These data are tabulated in Table G-2. An examination of these data leads to the same conclusions presented above. Thus, based on this analysis and not considering catalyst deterioration, it appears that an averaging time of 6 hours is suitable for determining SO_2 emission concentrations and that emissions levels established somewhat above commonly accepted vendor/contractor guarantees by 10 to 20 percent could be viewed as acceptable for purposes of allowing normal, short-term fluctuations.

G.2 CATALYST DETERIORATION

Due to the lack of substantial numerical qualification of the effect of catalyst deterioration on SO_2 emissions from sulfuric acid plants, SO_2 emission data gathered by simultaneous U.S. Environmental Protection Agency (EPA) source testing of the No. 6 and No. 7 plants at the Kennecott Garfield smelter during the period of June 13-16, 1972, were analyzed. The No. 6 (Parsons) plant began operating in February 1967 and was in the second month of its 12-month catalyst cleaning cycle during the source test. The No. 7 (Monsanto) plant began operation in September 1970 and was in the twelfth and last month of its catalyst cleaning cycle. The SO_2 emission data are tabulated in Table G-3.

A statistical analysis of these data leads to the conclusion that the 30-percent greater average emissions of the No. 7 plant, compared to the average emissions of the No. 6 plant, are statistically significant at the 90-percent probability level. It should be noted, however, that this difference in emissions reflects not only catalyst deterioration but other factors as well, such as a difference in emissions due to design or construction variations between Parsons 1967 acid plant technology and Monsanto 1970 acid plant technology. On the other

TABLE G-3. SUMMARY OF OUTLET SO_2 CONCENTRATIONS (ppm)

Run	No. 6 Plant	No. 7 Plant
2	389	296
3	753	855
4	1,036	2,277
5	1,745	1,207
6	938	1,131
7	1,608	2,553
8	794	1,104
9	1,128	1,355
10	930	1,433
Average	1,036	1,357

hand, it is probably safe to assume that the major portion of this difference in emissions is due to catalyst deterioration. Thus, the results of this analysis can be reviewed as indicating first, that catalyst deterioration does not have a significant effect on SO_2 emissions and second, that with a 12-month catalyst cleaning cycle, this difference in emissions due to deterioration appears to be of the order of magnitude of 30 percent.

G.3 ADDITIVE EFFECT OF EMISSION VARIATIONS AND CATALYST DETERIORATION

As discussed above, not considering catalyst deterioration, sulfuric acid plant performance standards based on 6-hour SO_2 emission levels 10 to 20 percent greater than commonly accepted vendor/contractor guarantees appear to be appropriate to allow short-term fluctuations in SO_2 emissions. As also discussed above, the increase in SO_2 emissions during the 12-month catalyst cleaning cycle can be estimated to be 30 percent. Based on the conservative assumption that catalyst deterioration is an increasing exponential function of time, almost all of the effect of catalyst deterioration will occur during the second half of the cleaning cycle. Because the emission variation data were based on the fifth month of the catalyst cleaning cycle, the data do not include significant catalyst deterioration and the increase in SO_2 emissions due to catalyst deterioration should be added to the allowance for new catalyst emission variation. Thus, considering short-term fluctuations of SO_2 emissions and using conservative assumptions regarding catalyst deterioration, new source performance standards (NSPS) can possibly be based upon 6-hour emission levels established 40 to 50 percent greater than commonly accepted vendor/contractor guarantees.

APPENDIX H

SULFUR DIOXIDE EMISSION TEST RESULTS FOR SINGLE-STAGE ABSORPTION SULFURIC ACID PLANTS PROCESSING METALLURGICAL OFFGAS STREAMS FROM PRIMARY COPPER SMELTERS

Please note: To provide the most comprehensive study possible, this appendix is reprinted, with minor editorial changes, from Volume I, Proposal Standards, of Background Information for New Soruce Performance Standards: Primary Copper, Lead, and Zinc Smelters, publication number EPA 450/2-74-002a.

APPENDIX H

SULFUR DIOXIDE EMISSION TEST RESULTS FOR SINGLE-STAGE ABSORPTION SULFURIC ACID PLANTS PROCESSING METALLURGICAL OFFGAS STREAMS FROM PRIMARY COPPER SMELTERS

H.1 BACKGROUND

Before emissions testing began in May 1972, the U.S. Environmental Protection Agency (EPA) surveyed all sulfur dioxide (SO_2) control systems at domestic primary copper smelters to determine which were most effective. Using the survey results, EPA selected for emission testing the facilities exhibiting the most advanced system design or highest degree of SO_2 emission reduction. The facilities selected consist of three single-stage absorption acid plants that treat offgasses from two different copper converting operations. All facilities were tested for SO_2 emissions using Reference Method 8 contained in Title 40 of the Code of Federal Regulations, Part 60 (40 CFR 60), Appendix A, first published in the Federal Register on December 23, 1971. Later, after one had been installed at a domestic copper smelter, a double-absorption acid plant was also tested. The analysis of this test is included in Appendix I.

During the initial portion of the testing program, the best domestic SO_2 control technology was considered to be single-stage absorption sulfuric acid plants (see Section 4.2). Thus, acid plants handling converter offgases had to be tested to determine the effects on acid plant performance of highly variable inlet SO_2 concentrations and flow rates.

All single-stage absorption acid plant tests were initally conducted using Method 8 of 40 CFR 60. However, to gain long-term operational data, an 8-week continuous monitoring test program was also conducted at one installation to monitor the frequently unsteady nature of converter offgas streams. The converter operation is a batch operation and, depending upon the number of converters in operation and their scheduling, will produce SO_2 concentrations and flow rates ranging from 0 percent to approximately 9 percent and flow rates ranging from 0 to the maximum blowing capacity of the converters.

Plant operating logs, acid plant inlet volumetric flow rate charts, absorber and converter temperature charts, and inlet concentration charts were reviewed to determine the operating condition of acid plants during the continuous monitoring program. Periods of startup and shutdown were eliminated from the data analysis, and the long-term SO_2 emission concentration averages were determined from the remaining valid data points. Finally, various averaging techniques were used to determine the most appropriate averaging interval, thereby masking the effect of massive short-term fluctuations.

H.2 SUMMARY OF TEST RESULTS

H.2.1 ASARCO--Hayden, Arizona

The copper converter single-absorption acid plant at the ASARCO smelter in Hayden, Arizona was tested during the week of June 19, 1972. The test consisted of eight separate runs using Reference Method 8 of 40 CFR 60. Two of the test runs were aborted because either the test equipment or the acid plant malfunctioned. Test 1 consisted of two samples, one for each orthogonal axis, whose results were combined to determine an overall emissions rate. In addition to the manual tests, continuous SO_2 monitoring was performed at the site for 2 days to provide comparative data experience for future tests. No statistical analysis of the continuous monitoring data was performed.

The ASARCO smelter has five copper converters, each requiring approximately 8 hours to process a batch of copper matte. The gas flow to the acid plant from the converters is as high as 2,830 Nm^3/min (100,000 scfm), depending upon the number of converters in operation. The gas stream to the acid plant has an SO_2 concentration of 4 to 9 percent.

The converter emissions are controlled by a 750-ton-per-day (tpd) single-absorption sulfuric acid plant designed by Chemiebau of West Germany and built in 1972 by Rust Engineering, U.S. Chemiebau's licensee. This acid plant was designed to process an inlet gas flow up to 2,830 Nm^3/min (100,000 scfm) at an SO_2 concentration of 4 percent. The acid plant has a four-stage capability, but only three catalytic stages were active during the test.

Table H-1 summarizes the results of the Hayden emission tests.

H.2.2 KENNECOTT--Garfield, Utah

The metallurgical, single-stage absorption sulfuric acid plants at the Kennecott smelter in Garfield, Utah, were tested during the week of June 19, 1972. A total of 20 acid mist and SO_2 emissions tests were conducted on two of the five acid plants. Specifically, Plants 6 and 7 were tested using Method 8 of 40 CFR 60, with 10 tests performed on each. Tables H-2 and H-3 summarize the manual emissions test results from the Kennecott-Garfield acid plants. In addition, a continuous SO_2 monitor was used to record long-term emissions from plant 7.

At the time of testing, 9 converters were in place at the Garfield facility. All offgases from these converters were ducted to six single-stage absorption sulfuric acid plants. Converter operations were scheduled to maintain a relatively constant SO_2 concentration in the acid plant feed streams. Each acid plant was designed to process a gas stream with an SO_2 concentration between 2 and 8 percent. The flow rate to each acid plant varied from 850 to 1,980 Nm^3/min (30,000 to 70,000 scfm), depending upon the number of converters in operation.

Acid plants 6 and 7 were chosen for the tests because they were then the newest installations at the facility. Plant 6, designed by Parsons Co., began operations in February 1967, was in the second month of its catalyst cleaning cycle during the test program, and is capable of processing up to 2,830 Nm^3/min (100,000 scfm) of gas at a concentration of 2 to 8 percent. Plant 7, designed by Monsanto Enviro-Chem and constructed by Leonard Construction Company, commenced operation in September 1970, was designed to handle the flow rate

TABLE H-1. SUMMARY OF EMISSION TEST DATA OBTAINED ASARCO-HAYDEN, JUNE 1972[a]

	Run number					
	1	2	3	4	5	Average
Date	June 20	June 20	June 21	June 21	June 22	
Test time (min)	145	144	144	145	144	144
Stack effluent						
Flow rate dscm/min (dscfm)	2,192 (78,300)	2,257 (80,600)	2,072 (74,000)	2,100 (75,000)	2,136 (76,300)	2,151 (75,770)
Temperature, °C (°F)	47 (116.00)	37 (99.00)	43 (110.00)	34 (93.00)	40 (104.00)	40 (104.00)
Pressure, mm Hg (in. Hg)	699 (27.5)	708 (27.87)	708 (27.87)	694 (27.33)	694 (27.33)	701 (27.58)
Acid plant SO_2 emissions						
ppm (by volume)	2,238	3,994	3,313	2,593	3,086	3,117
kg/dscm × 10^{-3}	29.1	51.9	43.0	22.9	40.1	37.4
(lb/dscf × 10^{-5})	(37.4)	(66.8)	(55.4)	(29.5)	(51.6)	(29.06)
kg/h (lb/h)	3,850.0 (1,750.0)	7,106.0 (3,230).0	5,411.5 (2,459.7)	2,920.5 (1,327.0)	5,197.0 (2,362.0)	4,896.6 (2,225.7)

[a] A single-stage absorption of Chemiebau design was tested. The plant processed copper converter offgases.

TABLE H-2. SUMMARY OF EMISSION TEST DATA OBTAINED AT THE NO. 6 (PARSONS) SINGLE-STAGE ABSORPTION SULFURIC ACID PLANT AT KENNECOTT-GARFIELD, JUNE 1972[a]

	Run number										
	1	2	3	4	5	6	7	8	9	10	Average
Date	June 13	June 14	June 14	June 14	June 15	June 15	June 15	June 16	June 16	June 16	
Test time (min)	112	56	56	112	112	112	112	112	112	112	101
Stack effluent											
Flow rate dscm/min (dscfm)	1,744 (62,800)	1,494 (53,300)	1,661 (59,800)	1,606 (57,900)	1,975 (71,100)	1,914 (68,900)	1,891 (68,100)	1,894 (68,200)	1,972 (71,000)	1,850 (66,600)	1,800 (64,804)
Temperature, °C (°F)	77 (169.0)	76 (167.0)	74 (165.0)	74 (164.0)	96 (203.0)	95 (196.0)	82 (181.0)	77 (169.0)	83 (182.0)	80 (175.0)	81 (178.0)
Pressure, mm Hg (in. Hg)	734 (28.90)	734 (28.90)	734 (28.90)	734 (28.90)	735 (28.92)	735 (28.92)	735 (28.92)	734 (28.90)	734 (28.90)	734 (28.90)	734 (29.91)
Concentration (SO_2)											
ppm (by volume)	126	388.5	752	1,036	1,744	938	1,608	7,940	1,128.0	930.0	944.7
kg/dscm × 10^{-4} (lb/dscf × 10^{-6})	16.3 (21)	50.5 (65)	97.1 (125)	134.0 (173)	227.0 (292)	122.0 (157)	209.0 (269)	103.0 (133)	146.9 (189)	930.0 (155)	122.6 (158)
kg/h (lb/h)	174.0 (79.1)	457.4 (207.9)	986.7 (448.5)	1,322.0 (601.0)	2,740.5 (1,245.7)	1,445.0 (657.0)	2,417.8 (1,099.0)	1,156.8 (544.0)	1,771.0 (805.0)	1,361.8 (619.0)	1,383.3 (628.7)

[a] The acid plant tested processed copper converter offgases.

TABLE H-3. SUMMARY OF EMISSION TEST DATA OBTAINED AT THE NO. 7 (MONSANTO) SINGLE-STAGE ABSORPTION SULFURIC ACID PLANT AT KENNECOTT-GARFIELD, JUNE 1972[a]

	Run number										
	1	2	3	4	5	6	7	8	9	10	Average
Date	June 13	June 14	June 14	June 14	June 15	June 15	June 15	June 16	June 16	June 16	
Test time (min)	111	56	56	112	112	112	112	112	112	112	100.7
Stack effluent											
Flow rate dscm/min (dscfm)	1,747.0 (62,900)	1,675.0 (60,300)	1,500.0 (54,000)	1,643.0 (59,150)	1,958.0 (70,500)	1,783.0 (64,200)	1,916.0 (69,000)	1,875.0 (67,500)	1,930.6 (69,500)	1,905.6 (68,600)	1,793.0 (64,560)
Temperature, °C (°F)	57 (135.0)	51 (124.0)	59 (138.0)	56 (134.0)	60 (139.0)	56 (133.0)	64 (146.0)	56 (134.0)	56 (134.0)	56 (134.0)	
Pressure, mm Hg (in. Hg)	734 (28.91)	734 (29.90)	734 (28.90)	734 (28.90)	735 (28.92)	735 (28.92)	735 (28.92)	734 (28.90)	734 (28.90)	734 (28.90)	734 (28.90)
Concentration (SO_2)											
ppm (by volume)	553	296	855	2,277	1,207	1,131	2,553	1,104	1,355	1,433	1,276
kg/dscm × 10^{-4} (lb/dscf × 10^{-6})	71.9 (92.5)	38.4 (49.5)	111.1 (143.0)	296.0 (381.0)	160.0 (202.0)	146.9 (189.0)	331.8 (427.0)	143.7 (185.0)	176.4 (227.0)	186.5 (240.0)	166.0 (213.6)
kg/h (lb/h)	768.0 (349.0)	393.8 (179.0)	1,019.0 (463.0)	2,974.8 (1,352.0)	1,879.8 (854.0)	1,601.7 (728.0)	3,889.0 (1,767.8)	1,648.0 (749.0)	2,082.5 (946.6)	2,173.0 (987.8)	1,842.7 (837.6)

[a] The acid plant tested processed copper converter offgases

fluctuations and SO_2 concentration associated with converter operations, and is capable of handling SO_2 concentrations ranging between 2 and 8 percent. Plant 7 was in the last month of its catalyst cleaning cycle when the manual tests were performed.

As noted earlier, a continuous monitoring test program was also conducted at the Kennecott-Garfield facility between September 15, 1972, and November 15, 1972, on Acid Plant 7 to gather long-term emissions data. The data being sought would be used to determine an averaging time that would effectively mask fluctuations in acid plant outlet concentrations and to evaluate the long-term performance capabilities of single-absorption acid plants. These emissions data were recorded by a Dupont 460 Continuous SO_2 Analyzer. Because Section 4.2 of this document discusses the results of that test, they are not discussed here.

APPENDIX I

ANALYSIS OF DUAL-ABSORPTION ACID PLANT CONTINUOUS SO_2 MONITORING DATA

Please note: To provide the most comprehensive study possible, this appendix is reprinted, with minor editorial changes, from Volume I, Proposal Standards, of Background Information for New Soruce Performance Standards: Primary Copper, Lead, and Zinc Smelters, publication number EPA 450/2-74-002a.

APPENDIX I

ANALYSIS OF DUAL-ABSORPTION ACID PLANT CONTINUOUS SO_2 MONITORING DATA

I.1 INTRODUCTION

The dual-absorption sulfuric acid plant at the ASARCO copper smelter at El Paso, Texas, was the first system of its type to be used in the domestic nonferrous smelting industry. The SO_2 emissions from this unit were measured by the U.S. Evironmental Protection Agency (EPA) beginning May 17, 1973, and continuing through December 14, 1973.

The objective of the test was to characterize the SO_2 emissions from a primary copper smelter using a control system of this type. The data were analyzed to determine the control system efficiency and any conditions which would cause high emissions. Finally, the emissions data were used to examine realistic and achievable SO_2 emission limitations for nonferrous smelting operations which produce strong SO_2 streams.

The ASARCO smelter at El Paso, Texas, is a custom copper smelter that produces 236 Mg/day (260 tons/day) of blister copper. Approximately 365 Mg/day (400 tons/day) SO_2 are also produced during the smelting process. The smelter operates three converters, with two converters operating at essentially all times while the third converter is in the pouring portion of its smelting cycle. This type of converter scheduling typically produces a relatively steady stream containing 3 to 7 percent SO_2.

The converter gases are controlled by the dual-absorption acid plant that produces approximately 450 Mg/day (500 tons/day) of sulfuric acid. The acid plant is designed to process a gas stream with an average inlet concentration of 4 percent, with an inlet concentration ranging between 2 percent to 10 percent SO_2, and an inlet flow rate of up to 2,830 Nm^3/min (100,000 cfm). The system is equipped with an automatic heater that permits efficient operation of the acid plant down to an inlet SO_2 concentration of approximately 2 percent. The catalyst renewal cycle of the acid plant is designed to be approximately once every 2 years.

The monitoring instrumentation included a Dupont 460 SO_2 analyzer for monitoring the outlet SO_2 concentration; a Beckman inlet SO_2 concentration analyzer; and a Westinghouse E2B 4-channel tape recorder, which permitted simultaneous recording of time, inlet SO_2 concentration, outlet SO_2 concentration, and inlet volumetric flow rate. The Beckman inlet SO_2 monitor was an integral part of the ASARCO SO_2 control system that required modification to permit recording of its output signal by the EPA recorders.

The accuracy of the outlet SO_2 monitoring instrumentation was verified as outlined in the proposed EPA Method 12 of 40 CFR 60. A total of nine manual Method 8 SO_2 tests, defined in 40 CFR 60, were performed between July 9 and 12, 1973. Table I-1 shows the results of the manual SO_2 measurements as determined by Method 8 and the corresponding SO_2 readings as determined by the Dupont 460 SO_2 monitoring instrument.

The entire monitoring program covered a period of 5,088 hours, or 212 days. During this time span, the acid plant was in operation for a total of 190 days, or 90 percent of the monitoring period. During the same time span, the monitoring instrumentation was in operation for 90 percent of the monitoring period. Including periods when both acid plant and monitoring instrumentation were inoperative, data were collected during 86 percent of the duration of the monitoring program. The monitoring instrumentation recorded one reading for each parameter monitored every 3 minutes. At the end of each 15-minute interval, an average of the previous five readings was computed. The 15-minute averages were used as the base data points for all subsequent computations and analyses.

TABLE I-1. COMPARISON OF SO_2 MEASUREMENTS USING EPA METHOD 8 AND THE DUPONT 460 SO_2 ANALYZER

	Test results (ppm SO_2)	
Date and time started	EPA Method 8	Dupont analyzer
7/09/73 (1617)	12.5	19.9
7/10/73 (1011)	122.0	121.2
7/10/73 (1418)	21.0	22.1
7/10/73 (1602)	117.5	116.3
7/10/73 (1745)	53.0	48.5
7/10/73 (0816)	19.5	22.2
7/10/73 (1000)	49.5	51.4
7/10/73 (1627)	239.0	224.3
7/10/73 (1805)	22.5	23.1

I.2 VALIDATION OF DATA

To ensure that the recorded data were representative of "normal" operating conditions, data validation criteria were established. The acid plant operations log, the acid plant engineer's log, the catalyst temperature charts, and the copper converter operating logs were reviewed to determine the operating state of the converter operations and the acid plant. Periods during which the acid plant was not operating and periods of excess emissions during startup were removed from the compiled data. For purposes of analysis of the compiled data, all other operating situations were considered normal.

During the the test program, the acid plant experienced a number of shutdown and startup situations. The periods of acid plant downtime lasted for as little as 30 minutes to as long as 5 days. A general review of the data showed that the shorter durations of downtime produced shorter periods of high emissions after startup than did the downtimes of longer duration. Therefore, each period of downtime and startup was evaluated to derive a quantitative relationship between the duration of the downtime and the duration of excess emissions after startup.

In developing an approximate relationship between the duration of abnormal emissions and the duration of downtime, a family of curves was prepared to show average emission vs. time after startup based on the data monitored. Figure I-1 shows the relationship between the downtime duration and the emissions rate immediately after startup. There were 25 startups during the monitoring period. These were categorizied into five groups depending upon downtime duration. The curves represent the following downtime periods: 1.99 hours or less, 2 to 5.99 hours, 6 to 9.99 hours, 10 to 13.99 hours, and greater than or equal to 14 hours. Each curve represents the following total number of downtimes: 7 downtimes of 1.99 hours or less, 3 downtimes of from 2 to 5.99 hours duration, 3 downtimes of from 6 to 9.99 hours duration, 4 downtimes of from 10 to 13.99 hours duration, and 7 downtimes of 14 hours or greater duration. Normal operation was considered attained when the average emissions decreased to 500 ppm.

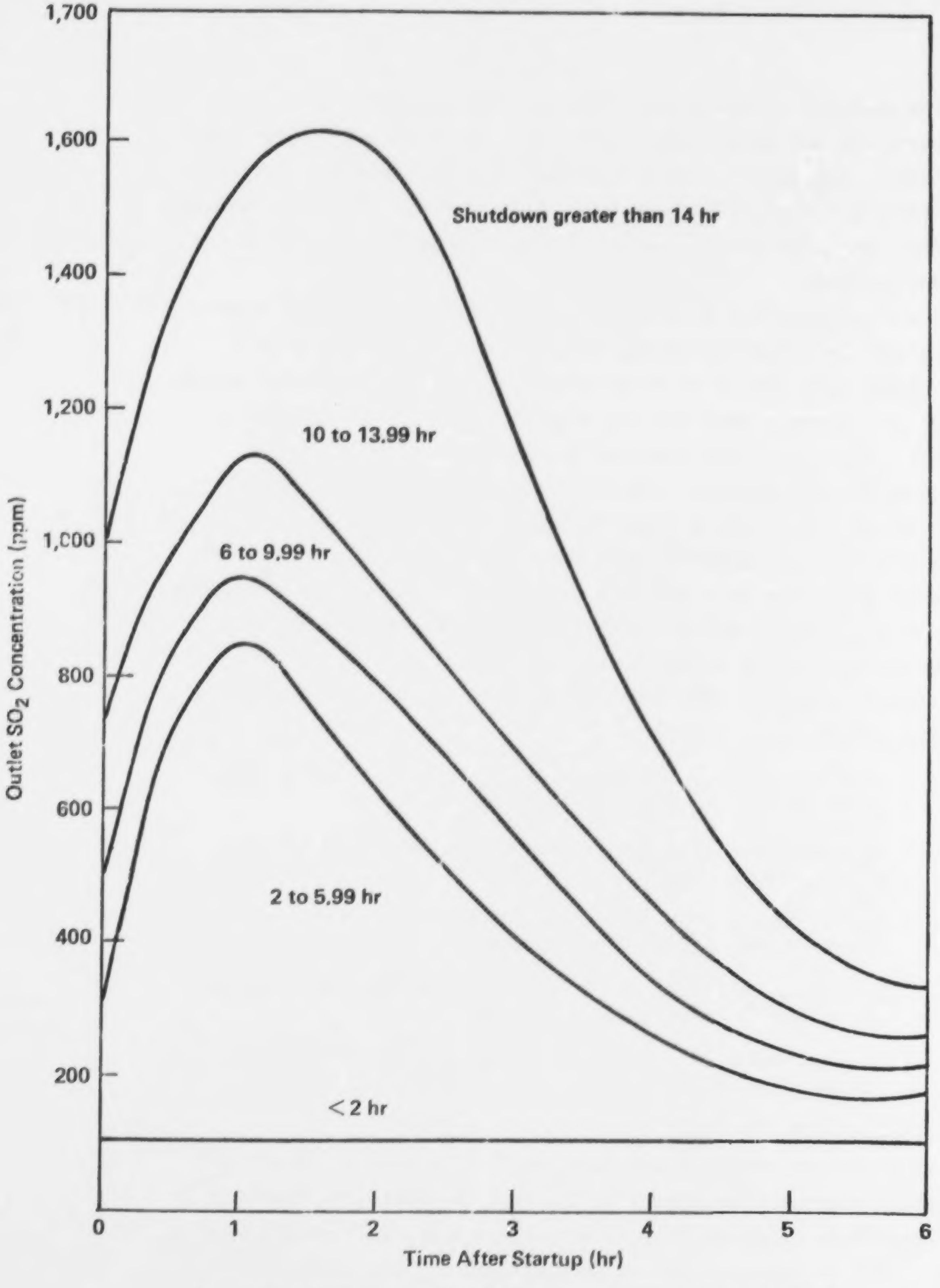

Figure I-1. Average emissions after startup versus time after startup.

The analysis of the curves indicates that downtimes of up to 1.99 hours did not cause excess emissions. Downtimes of greater than 14.99 hours, however, typically resulted in excess emissions for up to approximately 5 hours after startup. Other shutdown intervals resulted in normal operation after a period of time ranging between the two previous extremes.

The exact duration of excess emissions during startup will vary because the time required to attain normal operation depends to a major degree upon the skill of the acid plant operator, his/her perception of the system's imbalance and his/her response with corrective measures. Also, the time required to attain normal operation is dependent upon the response time of the acid plant process control system to any corrective actions initiated by the operator. The curves of Figure I-1 indicate that there may be considerable elapsed time after startup before the acid plant regains equilibrium conditions. Based on the curves, data validation criteria were developed for startup periods. Data points during the initial portions of an acid plant startup were excluded from the analysis based on the following criteria, to the nearest hours:

- For shutdowns of less than 2 hours, the first valid datum point occurs immediately after startup.
- For shutdowns of 2 to 5.99 hours, the first valid datum point occurs 3 hours after startup.
- For shutdowns of 6 to 9.99 hours, the first valid datum point occurs 4 hours after startup.
- For shutdowns of 10 to 13.99 hours, the first valid datum point occurs 4 hours after startup.
- For shutdowns of greater than 14 hours, the first valid datum point occurs 5 hours after startup.

I.3 DISCUSSION OF THE DATA

With periods of acid plant downtime and the initial portion of acid plant startup eliminated from the recorded data, the remaining data constitute emissions from normal smelting and acid plant operations. This includes periods of abnormally low inlet concentration when all converters were out of the hoods for short periods. These situations are common occurrences in copper converter operations.

As previously discussed, the inlet SO_2 concentration to the acid plant was measured at 3-minute intervals. The readings were then averaged every 15 minutes to determine the 15-minute average base data points. The inlet gas stream averaged 3.80 percent SO_2 for the entire test period, with a standard deviation of 1.64 percent SO_2. The highest recorded 15-minute average inlet for the total monitoring period was 9.19 percent SO_2.

An analysis of the distribution of the 15-minute inlet SO_2 readings indicated that the acid plant processed gases of greater than 3.5 percent for only approximately 55 percent of the time. Figures I-2 and I-3 show the concentration distribution and the cumulative frequency distribution of the inlet gas stream SO_2 concentrations recorded during the monitoring period.

I.3.1 Catalyst Deterioration

The efficient operation of any acid plant is governed to a major degree by the condition of the catalyst that aids the conversion reaction of SO_2 to SO_3. As the catalyst is used, its condition can deteriorate and thus decrease the control efficiency of the system. This naturally results in increased emissions from the acid plant. To ascertain any change in conversion efficiency attributable to catalyst use, the change in efficiency was determined for various time intervals over the total test period. The implied assumption in this procedure was that any decrease in control efficiency would be basically due to the decreased reactivity of the catalyst.

The acid plant conversion efficiency was calculated using the following definition:

$$\text{Efficiency, } E = \frac{\text{Mass } SO_2 \text{ converted}}{\text{Mass } SO_2 \text{ available}} .$$

Adopting the ideal gas law for SO_2, the previous definition can be represented by the equation:

$$E = \left(1 - \frac{C_{out}}{C_{in}}\right) \left(1 + C_{out} + C^2{}_{out} + \ldots + C^n{}_{out}\right)$$

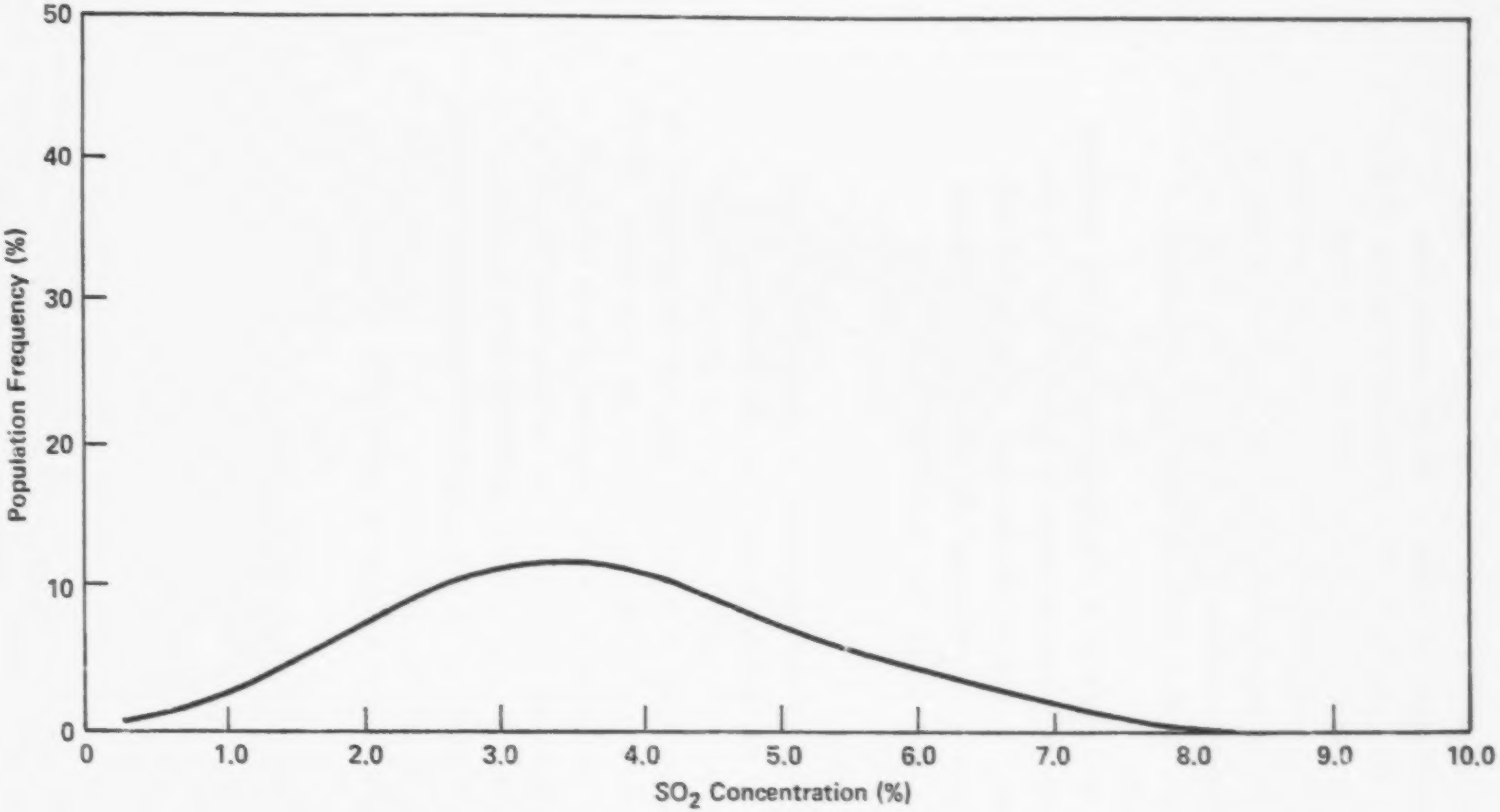

Figure I-2. Inlet SO_2 concentration frequency distribution.

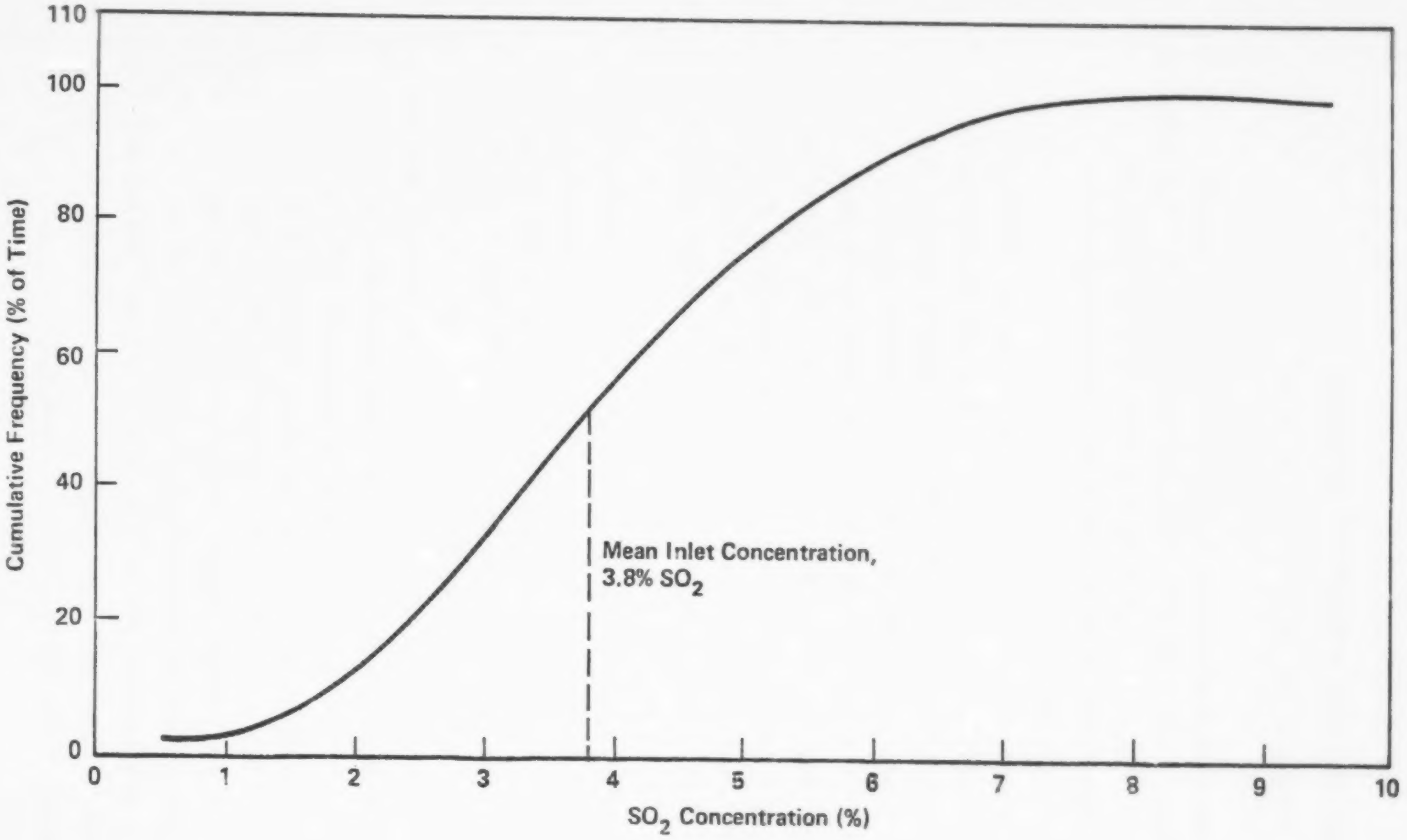

Figure I-3. Inlet SO_2 concentration cumulative frequency distribution.

where

C_{in} = SO_2 concentration entering the acid plant

C_{out} = SO_2 concentration leaving the acid plant.

The acid plant commenced operation in December 1972. Between May 1973 and December 1973, the acid plant was monitored while operating for approximately 171 days, or approximately 86 percent of the time. At the end of the monitoring program, the acid plant has been in operation a total of 335 days.

The normal cleaning cycle for the acid plant catalyst, based on the manufacturer's design, is 2 years. Thus, the system was monitored during the second quarter of its normal catalyst cleaning cycle. Due to the failure of parts of the gas precleaning system to operate properly, however, the catalyst deterioration rate was accelerated, and the acid plant catalyst was screened during March 1974. Based on this information, the catalyst renewal cycle therefore covered a period of 1.2 years, and the acid plant was considered to have been monitored during the second and third quarters of its catalyst cleaning cycle.

One least-squares regression analysis of the change in efficiency with usage covers the total test period from May 17, 1973, through December 14, 1973. Similarly, second and third analyses of the change in efficiency with time were also made and included the last 2 months and the last month of the monitoring period, respectively. A review of the three results indicates that the acid plant's efficiency remained essentially constant at an average of greater than 99.70 percent during the total test program. The respective changes in efficiency within the observed periods indicated by the three analyses were -0.20×10^{-7}, -5.6×10^{-7}, and -8.7×10^{-7} percent per day. The minimum efficiencies from these changes in efficiency were 99.750 percent, 99.643 percent and 99.688 percent, respectively. Thus, neither within a given interval nor between one reporting interval and other did the analysis show sufficient changes in efficiency to indicate a significant change in the condition of the catalyst.

I.3.2 Effect of Inlet SO_2 Concentration on Emissions

The most important aspect of the inlet SO_2 concentration is its effect on acid plant operating efficiency and the resulting outlet SO_2 concentration. To ascertain the effects of varying inlet SO_2 concentrations on the resulting outlet SO_2 concentrations, all of the simultaneous 15-minute inlet and outlet concentration data were used to develop a least-squares straight line. The results of this analysis indicated there is a direct linear relationship between inlet SO_2 concentration and the resulting outlet SO_2 concentration. The correlation coefficient of the analysis was calculated to be 0.413 and was determined to be significant enought to warrant a conclusion of linearity. Figure I-4 shows the graph of the least-squares line and its standard error.

The inlet SO_2 concentrations experienced during this test were somewhat lower than the concentrations of 5 to 6 percent achievable from typical copper converter operations. With an average of 3.8 percent SO_2 and a standard deviation of 1.64 percent SO_2, approximately 68 percent of the readings were between 2.2 and 5.4 percent SO_2, indicating that the inlet concentrations are biased low and thus result in lower outlet concentrations. The fact that the acid plant inlet concentration was typically low indicates that the typical outlet concentration was lower than that expected from other similar acid plants operating at a higher average inlet concentration. This factor must be taken into account when determining emissions limits for other smelting operations, based on data from this test.

An inlet concentration of 9 percent is approximately the maximum inlet SO_2 concentration that can be processed by most modern dual-stage acid plants. Figure I-4 is significant, therefore, when predicting the expected emissions from a smelter generating an inlet gas stream within the observed range of this test (0.02 to 9.16 percent SO_2). It shows that the average outlet concentration increases approximately 50 ppm per 1 percent increase in inlet concentration above 3.8 percent. For instance, when the average inlet concentration to the acid plant was 9 percent SO_2, the average emission rate indicated from the test

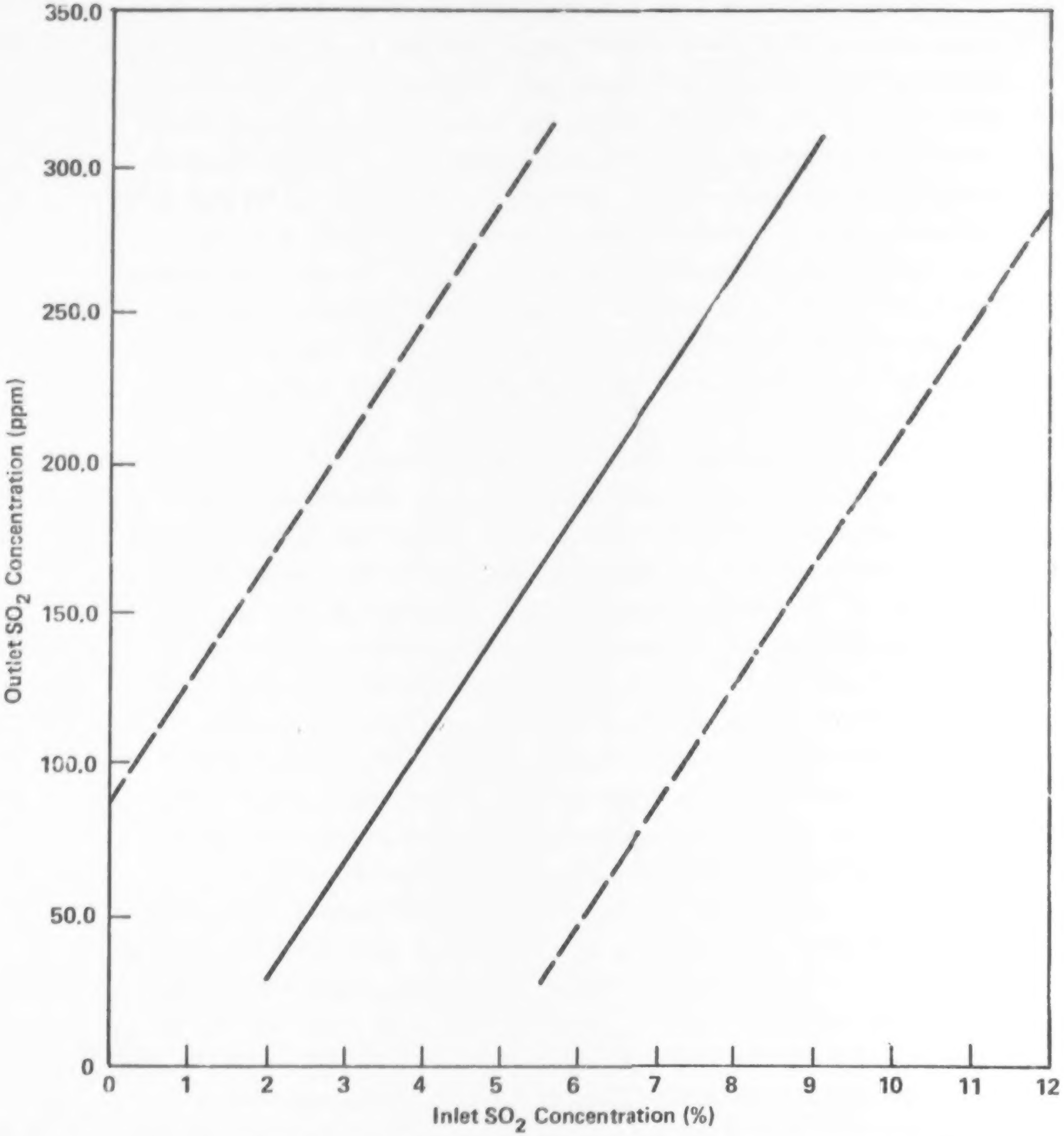

Figure I-4. Outlet SO_2 concentration versus inlet SO_2 concentration.

was approximately three times the emission rate obtained at 3.8 percent inlet SO_2. This increase is basically due to increased inlet concentration at a constant conversion efficiency.

I.4 RESULTS OF THE TEST PROGRAM

The results of the test program indicated that during normal operations the average emissions, based on 15-minute readings, ranged between 10 and 2,920 ppm. Approximately 90 percent of these values, however, were below 250 ppm and well below the typical manufacturer's guaranteed emission rate of 500 ppm.

There were, however, periods of relatively high emissions, even when averaged over 6-hour periods, which could not be attributed to malfunctions, startups, or shutdowns. It was thought that these periods might be caused by relatively high inlet concentrations, resulting in a corresponding increase in outlet concentrations. To examine this possibility, 6-hour averages of 400 ppm or greater were located in the data base, and the 24 15-minute inlet concentration readings that made up the 6-hour averages were recorded. The concentration frequency distributions of these inlet readings were then compared with the inlet concentration frequency distribution for the entire monitoring period. In general, the individual distributions did not vary significantly enough from the composite for the entire monitoring period to indicate that the excursions occurred during periods of unusually high or during abnormal inlet concentration conditions. The catalyst converter temperatures and inlet gas flow rates were also reviewed, but no abnormalities were noted in the parameters.

Because the periods of relatively high emissions were not caused by abnormal inlet gas conditions or by abnormal operation of the acid plant system, the compiled data were averaged over various time intervals ranging from 1 to 10 hours to examine the effect of averaging time on damping of normal excursions. As a result, the effects of normal short-term excursions were spread over successively longer periods of time. Table I-2 shows a matrix, to the nearest 0.05

TABLE I-2. THE EFFECT OF REFERENCE CONCENTRATION LEVEL AND AVERAGING TIME ON THE PERCENTAGE OF EXCURSIONS

Averaging time	Number of readings	Reference concentration level, ppm													Maximum concentration, ppm
		150	200	250	300	350	400	450	500	550	600	650	700	750	
15 min	14,612	20.00	15.00	10.00	7.50	5.00	4.00	3.00	2.30	1.60	1.35	1.15	1.05	1.05	2,920
1 hr	3,628	20.00	15.00	10.00	7.10	4.10	3.15	2.65	2.10	1.75	1.40	1.00	0.90	0.80	1,582
2 hr	3,702	20.00	15.00	10.00	5.00	3.00	2.50	2.00	1.75	1.50	1.25	1.00	0.90	0.70	1,261
3 hr	3,758	20.00	15.00	10.00	5.00	2.20	2.00	1.60	1.25	0.85	0.80	0.55	0.50	0.50	1,238
4 hr	3,803	20.00	8.15	6.10	3.00	2.20	1.40	1.05	0.80	0.75	0.50	0.45	0.30	0.25	935
5 hr	3,841	20.00	10.00	5.00	2.75	1.75	1.25	1.00	0.75	0.55	0.40	0.30	0.25	0.15	935
6 hr	3,876	20.00	10.00	5.00	2.45	1.75	1.20	0.90	0.45	0.35	0.30	0.15	0.05	0.05	752
7 hr	3,907	20.00	10.00	5.00	2.15	1.40	1.00	0.55	0.30	0.20	0.10	0.05	0.00	0.00	662
8 hr	3,935	15.00	10.00	5.00	2.15	1.40	0.80	0.50	0.25	0.10	0.05	0.00	0.00	0.00	662
10 hr	3,988	15.00	10.00	5.00	2.05	1.20	0.55	0.25	0.10	0.05	0.00	0.00	0.00	0.00	576

percent, of the percentages of the total readings that exceeded given concentrations for various averaging intervals.

It can be seen from Table I-2 that, as the averaging time for a given concentration level increases, the percentage of excursions above that concentration level tends to converge to zero. For example, Table I-2 indicates that from 20 to 15 percent of the recorded values exceeded 150 ppm, depending on the averaging intervals between 1 and 10 hours.

Similarly, in Table I-2, an increase in the concentration level for a given averaging time will also cause the matrix to converge rapidly to a small value. For example, observing the 6-hour averaging interval, there is a 20 percent excursion rate at the 150 ppm level. Increasing the concentration level to 300 ppm decreases the excursion rate to 2.45 percent; increasing the concentration level to 750 ppm decreases the excursion rate to 0.05 percent.

Based on the results of Table I-2, as either the averaging time increases, the concentration level increases, or both increase, the percentage of excursions tends to converge toward a small value in the matrix.

I.5 CONCLUSIONS

As previously indicated, the typical manufacturer's guarantee for a dual-stage acid plant is 500 ppm, based on a 5 to 6 percent average inlet SO_2 concentration. The results of the test, however, indicated that the test was carried out at a 3.8 percent average inlet concentration, somewhat lower than the average inlet concentration from typical copper converting operations. The test results also indicate that there is a direct linear relationship between inlet gas-stream SO_2 concentration and outlet gas-stream SO_2 concentration; the inlet concentration increases in proportion to the outlet concentrations. Therefore, because the inlet concentration was somewhat lower than normal, the resulting outlet concentration was considered lower than that from typical copper smelters.

Because the manufacturer's guarantee of 500 ppm is based on a 5 to 6 percent inlet SO_2 concentration into a typical smelter converter acid plant, the equivalent SO_2 concentration for the ASARCO acid plant during the test period was 400 ppm. This is due to the typically lower inlet concentrations.

As discussed in Appendix H, an appropriate averaging time for masking outlet concentration fluctuations from single-stage absorption acid plants was determined to be 6 hours. The test of the ASARCO plant indicates that a 6-hour averaging time is also sufficient to mask fluctuations from a dual-absorption acid plant. The results show that an emission rate of 400 ppm for a 6-hour averaging time would result in 1.20 percent excursions.

Although the results of this test program indicate that a reasonable emissions limit equivalent to the vendor's guarantee (400 ppm) would result in only 1.20 percent violation rate, the effects of higher inlet SO_2 concentrations at other smelting operations and acid plant catalyst deterioration must be taken into account. To account for situations of increased emissions due to higher inlet concentrations of up to 9 percent, the results of Table I-3 require prorating upward a maximum of 200 ppm.

The results of this test were not conclusive as to the characteristics of increased emissions due to catalyst deterioration because no deterioration was observed during this test. Discussions with the designers of the ASARCO acid plant indicated that up to a 10-percent increase in emissions was expected before renewal of the catalyst. This factor, therefore, has to be taken into account when predicting the expected emissions from a system of this type. Based on the previous factors, the results of Table I-3 were prorated upward to take higher inlet concentrations and catalyst deterioration into account.

Table I-3 shows an acid plant operating at an inlet of as high as 9 percent and taking catalyst deterioration into account. From Table I-3 it can be seen that an acid plant processing the maximum expected inlet concentration could be expected to maintain an emission rate of 650 ppm with only a 1.20 percent excursion rate.

TABLE I-3. PRORATED EXCURSION PERCENTAGES[a]

	Outlet SO_2 concentration, ppm													Expected concentration, ppm
	150	200	250	300	350	400	450	500	550	600	650	700	750	
Percentage of averages exceeding outlet SO_2 concentration	--	--	--	--	--	20.00	10.00	5.00	2.45	1.75	1.20	0.90	0.45	977

[a] Based upon a 6-hour averaging time and a total of 3,876 readings.

In general, however, a new source performance standard (NSPS) set at the 650-ppm level and a 6-hour averaging time would result in a probable excursion rate of less than 1.20 percent. The general NSPS provisions (39 FR 9308) specify that each performance test for the purpose of compliance shall consist of the arithmetic mean of the results from three separate runs. To determine the number of times that the ASARCO acid plant exceeded the 400-ppm level (equivalent to 650 ppm in Table I-2), the recorded data from the test program were reviewed. Each 6-hour average of 400 ppm or greater was considered an excursion. Readings for 24 hours both before and after the violation were reviewed to determine whether the average of any two readings together with the excursion would exceed 400 ppm. The three 6-hour averaging periods were chosen so that none of the periods overlapped. The results indicate that, of 48 recorded readings greater than 400 ppm during the entire monitoring period, only 6 result in averages of 3 runs greater than 400 ppm. From this evaluation, the probable percentage of 6-hour averages in excess of 650 ppm, based on a 9 percent SO_2 inlet stream, would be approximately 0.15 percent.

APPENDIX J

EXAMPLE CALCULATIONS
MODEL PLANT OPERATING PARAMETERS

This appendix contains examples of calculations used in the development of the Background Information Document for the review of the primary copper smelter NSPS. Calculations are for Control Alternative I-G, new greenfield smelter processing high-impurity materials (oxyfuel burners and 100 percent blending of the reverberatory furnace offgas stream). Where procedures for the expansion scenarios differ, example calculations are included for selected expansion scenarios.

The following input data obtained from Chapter 6 are reproduced for the reader's convenience.

INPUT DATA

	New greenfield smelter	Expansion Scenario 9
Feed:		
Mg/day	1,364	2,045
Copper (%)	22.9	22.4
Sulfur (%)	27.0	28.4
Iron (%)	19.6	24.8
Sulfur removal (%)		
Roaster	19.3	-
Reverberatory furnace	28.4	41.2
Converter	52.3	59.8
Matte grade (%)	40.0	39.0

The assumptions used in these calculations are those listed on pages 6-6 and 6-9. Additional assumptions are indicated in the example calculations.

J.2 NEW GREENFIELD SMELTER PROCESSING HIGH-IMPURITY MATERIALS

J.2.1 Material balance:

Copper in feed: $1{,}364 \times 0.229 = 312.4$ Mg/day

Matte fall: $312.4 \div 0.40 = 781.0$ Mg/day

Inert in matte: $781.0 \times 0.10 = 78.1$ Mg/day

S as CO_2S: $312.4 \times \frac{32}{127.1} = 78.7$ Mg/day

CO_2S: $312.4 + 78.7 = 391.1$ Mg/day

FeS: $781.0 - 78.1 - 391.1 = 311.8$ Mg/day

S as FeS: $311.8 \times \frac{32}{87.8} = 113.6$ Mg/day

S in matte: $113.6 + 78.7 = 192.3$ Mg/day

S in feed: $1{,}364 \times 0.270 = 368.3$ Mg/day

S removed in MHR and RV: $368.3 - 192.3 = 176.0$ Mg/day

S removed in MHR: $176.0 \times \frac{0.193}{0.193 + 0.284} = 71.2$ Mg/day

S removed in RV: $176.0 \times \frac{0.284}{0.193 + 0.284} = 104.8$ Mg/day

J.2.2 Multihearth Roaster

Volumetric flow: 828.5 Nm^3/min [at 70° F, 1 atm]

$$\frac{71.2 \text{ Mg S removed}}{\text{day}} \times \frac{1 \text{ day}}{1{,}440 \text{ min}} \times \frac{\text{Mg}\cdot\text{mol}}{32 \text{ Mg}} \times \frac{22.4 \times 10^3 \text{ Nm}^3}{\text{Mg}\cdot\text{mol}}$$

$$\times \frac{\text{Nm}^3 \text{ offgas}}{0.045 \text{ Nm}^3 \text{ SO}_2} \times \frac{530}{492} = 828.5$$

Fraction O_2: 0.165

Theoretical air: $\frac{828.5 \times 0.045}{0.21} = 177.5$ Nm^3/min

Dilution air: $828.5 - 177.5 = 651.0$ Nm^3/min

Fraction O_2: $\frac{651.0 \times 0.21}{828.5} = 0.165$

J.2.3 Reverberatory Furnace

Natural gas equivalent required:[a,b] 81.3 Nm^3/min [70° F, 1 atm]

$$\frac{(1{,}364.0 - 71.2)\ \text{Mg calcine}}{\text{day}} \times \frac{1\ \text{day}}{1{,}440\ \text{min}} \times \frac{1.1\ \text{tons}}{\text{Mg}}$$

$$\times \frac{(0.6 \times 4.5 \times 10^6)\ \text{BTO}}{\text{Ton feed}} \times \frac{\text{ft}^3\ \text{natural gas}}{1{,}000\ \text{Btu}}$$

$$\times \frac{(2.832 \times 10^{-2}\ \text{m}^3)}{\text{ft}^3} \times \frac{530}{492} = 81.3$$

Combustion products ($CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$):

CO_2: = 81.3 Nm^3/min

H_2O: = 162.6 Nm^3/min

N_2: $162.6 \times \frac{0.79}{0.21}$ = 611.7 Nm^3/min

SO_2 formed: 54.9 Nm^3/min [79° F, 1 atm]

$$\frac{104.8\ \text{Mg S removed}}{\text{day}} \times \frac{1\ \text{day}}{1{,}440\ \text{min}} \times \frac{\text{Mg}\cdot\text{mol}}{32\ \text{Mg}}$$

$$\times \frac{(22.4 \times 10^3\ \text{Nm}^3)}{\text{Mg}\cdot\text{mol}} \times \frac{530}{492} = 54.9$$

N_2 in air to form SO_2: $54.9 \times \frac{0.79}{0.21}$ = 206.5 Nm^3/min

Air leakage: 511.3 Nm^3/min

O_2 requirements: $162.6 + 54.9$ = 217.5 Nm^3/min

N_2 at 65/35 H_2/O_2: $217.5 \times \frac{0.65}{0.35}$ = 403.9 Nm^3/min

---

[a] Natural gas equivalents are used in reverberatory furnace calculations for convenience. It is assumed that combustion products per Btu of input from natural gas are essentially the same as combustion products per Btu of input from other fossil fuels.

[b] The literature indicates a 40-percent reduction in heat requirements when using oxyfuel burners.

Air leakage: 403.9 ÷ 0.79 = 511.3 Nm^3/min

Moisture in air (70° F 40% RH): = 3.9

511.3 × 0.00757 = 3.9 Nm^3/min

Theoretical RV offgas: = 706.6 Nm^3/min

Component	Nm^3/min
CO_2	81.3
H_2O	166.5
Combustion	(162.6)
In air	(3.9)
N_2	403.9
SO_2	54.9

Dry 65/35 air to result in 1 percent O_2 at offtake: 20.8 Nm^3/min

$$\frac{0.35 \text{ volume air}}{\text{Volume air} + 706.6} = 0.01$$

N_2 in makeup air: 20.8 × 0.65 = 13.5 Nm^3/min

O_2 in makeup air: 20.8 × 0.35 = 7.3 Nm^3/min

H_2O in makeup air: 20.8 × 0.00757 = 0.2 Nm^3/min

RV offgas at offtake:

	Dry basis		Wet basis	
Component	Nm^3/min	%	Nm^3/min	%
CO_2	81.3	14.5	81.3	11.2
H_2O	-		166.7	22.9
(combustion)			(162.6)	
(leakage air)			(3.9)	
(makeup air)			(0.2)	
N_2	417.4	74.4	417.4	57.4
(leakage air)	(403.9)		(403.9)	
(makeup air)	(13.5)		(13.5)	
SO_2	54.9	9.8	54.9	7.5
O_2	7.3	1.3	7.3	1.0
Total	560.9		727.6	

Leakage through waste heat boiler and ESP: = 727.6 Nm^3/min

RV gas to acid plant (dry basis): = 1,284.4 Nm^3/min

Component	Nm^3/min	%
CO_2	81.3	6.3
N_2	992.2	77.0
(at offtake)	(417.4)	
(leakage)	(574.8)	
O_2	160.1	12.4
(at offtake)	(7.3)	
(leakage)	(152.8)	
SO_2	54.9	4.3
	1,288.5	

J.2.4 Converters

J.2.4.1 General. The availability of a strong SO_2 stream can significantly enhance the attractiveness of weak-stream blending as a means by which to control weak SO_2 streams. Consequently, a converter scheduling scheme that will maximize the converter offgas SO_2 concentration over time is highly desirable.

A typical converter cycle can take between 11 and 12 hours per charge. In converting a 40-percent matte, a slag blow will last approximately 6 hours, while a copper blow lasts about 3 hours. Time taken up by charging, pouring, and skimming will generally be about 2 hours per cycle. Figure J-1 presents the converter schedule used in this study to determine the time profile of the total converter offgas flow for a 40-percent matte. Copper and slag blowing fumes will change when other matte grades constitute the converter charge. A three-converter operation performing five converter cycles per 24-hour period was chosen as representative of domestic practice. An intercycle time of 3-2/3 hours was determined to be typical. Offgas profile of the converter aisle is determined in the following paragraphs.

J.2.4.2 Slag Blow. [$FeS + 1.5\ O_2 = FeO + SO_2$]

FeS: = 311.8 Mg/day

FeS/cycle: 311.8 ÷ 5 = 62.4 Mg/cycle

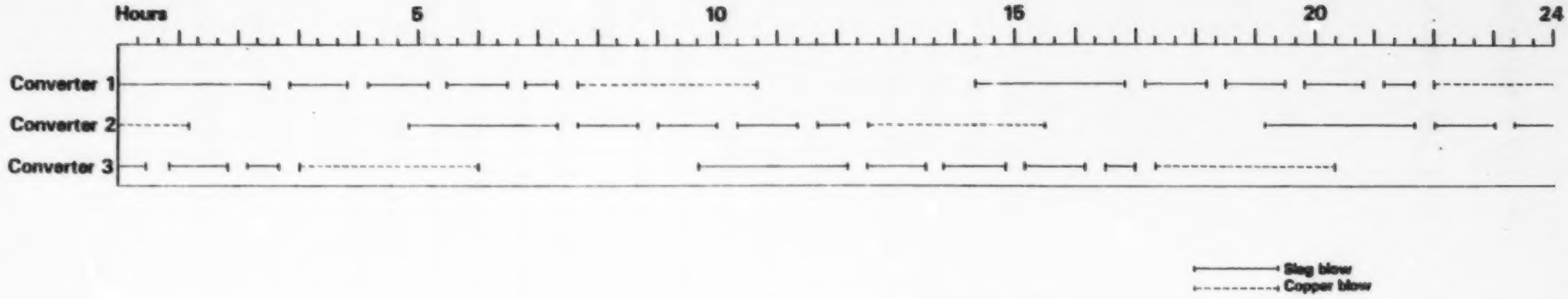

Figure J-1. Model smelter converter operating schedule.

SO_2: = 47.6 Nm^3/min
[70° F, 1 atm]

$$\frac{62.4\ \text{Mg FeS}}{\text{cycle}} \times \frac{1\ \text{cycle}}{360\ \text{min}} \times \frac{\text{Mg}\cdot\text{mol}}{87.8\ \text{Mg FeS}} \times \frac{(22.4 \times 10^3\ \text{Nm}^3)}{\text{Mg}\cdot\text{mol}} \times \frac{530}{462} = 47.6$$

Theoretical O_2: 47.6×1.5 = 71.4 Nm^3/min

Actual O_2: $71.4 \div 0.75$ = 95.2 Nm^3/min

Actual N_2: $95.2 \times \frac{0.79}{0.21}$ = 358.1 Nm^3/min

Offgas before dilution: = 429.5 Nm^3/min

$47.6 + 95.2 - 71.4 + 358.1 = 429.5$

Offgas after dilution: 2×429.5 = 859.0 Nm^3/min

Fraction SO_2: $47.6 \div 859.0$ = 0.0554

Fraction O_2: = 0.133

$$\frac{95.2 - 71.4 + 0.21 \times 429.5}{859.0} = 0.133$$

J.2.4.3 Copper Blow [$Cu_2S + O_2 \rightarrow 2\ Cu + SO_2$].

Cu_2S: = 391.1 Mg/day

Cu_2S/cycle: $391.1 \div 5$ = 78.2 Mg/cycle

SO_2: = 65.9 Nm^3/min

$$\frac{78.2 \times 22.4 \times 10^3 \times 530}{180 \times 159.1 \times 492} = 65.9$$

Actual O_2: $65.9 \div 0.75$ = 87.9 Nm^3/min

Actual N_2: $87.9 \times \frac{0.79}{0.21}$ = 330.7 Nm^3/min

Offgas before dilution: $330.7 + 87.9$ = 418.6 Nm^3/min

Offgas after dilution: 2×418.6 = 837.2 Nm^3/min

Fraction SO_2: $65.9 \div 837.2$ = 0.0787

Fraction O_2: = 0.131

$$\frac{87.9 - 65.9 + 0.21 \times 418.6}{837.2} = 0.131$$

J.2.4.4 Offgas Profile. Using the converter aisle station presented in Figure 6.2 and the slag and copper blow flows determined above the following converter aisle offgas profile can be developed.

Converter aisle status						
Number of converters						
Slag blow	Copper blow	Off-stack	Hr/day	Nm^3/min	SO_2 (%)	O_2 (%)
2	1	0	4.0[a]	2,555.2	6.30	13.2
2	0	1	5.2	1,718.0	5.54	13.3
1	1	1	9.5	1,696.2	6.69	13.2
1	0	2	2.3	859.0	5.54	13.3
0	1	2	1.7	837.2	7.57	13.1
0	0	3	1.3	0	-	-
Average flow				1,611.1	6.31	13.2

---

[a] Example calculation:

Flow: $2 \times 859.0 + 837.2 = 2{,}555.2$ Nm^3/min

Fraction SO_2: $$\frac{2 \times 859.0 \times 0.0554 + 837.2 \times 0.0787}{2{,}555.2} = 0.063$$

Average flow: $$\frac{\Sigma \text{ hr pending} \times Nm^3/\text{min}}{24}$$

J.2.5 Acid Plant Flows

J.2.5.1 Feed. Blended MHR, RV, and CV streams are fed to the acid plant. Profile of this blended stream is determined as shown below:

Stream	Hr/day	Nm^3/min	SO_2 (%)	O_2 (%)
Multihearth roaster	24	828	4.50	16.5
Reverberatory furnace	24	1,284	4.30	12.4
Converter aisle	4.0	2,555	6.30	13.2
	5.2	1,718	5.54	13.3
	9.5	1,696	6.69	13.2
	2.3	859	5.54	13.3
	1.7	837	7.87	13.1
	1.3	0	-	-
To acid plant	4.0[a]	4,667	5.43	13.6
	5.2	3,830	4.90	13.7
	9.5	3,808	5.41	13.7
	2.3	2,971	4.71	13.8
	1.7	2,949	5.37	13.8
	1.3	2,112	4.36	14.0
	Average	3,723	5.21	13.7

---

[a] Example calculation (see next page):

Flow: $828 + 1,284 + 2,555 = 4,667\ \text{Nm}^3/\text{min}$

Fraction SO_2:

$$\frac{828 \times 0.045 + 1,284 \times 0.043 + 2,555 \times 0.063}{4,667} = 0.0543$$

J.2.5.2 Effluent. Acid plant effluent is based on average flows, 98.3 percent conversion efficiency, and on the assumption that only SO_2 is removed from the dry gas in the acid plant. The following model was developed for this purpose.

V: Volume of dry gas to the acid plant

S: Fraction of SO_2 in dry inlet gas

O: Fraction O_2 in dry inlet gas.

SO_2 converted: $0.983\ VS$

O_2 used: $\frac{0.983\ VS}{2}$

$$SO_2 + \tfrac{1}{2}O_2 + H_2O \rightarrow H_2SO_4$$

Effluent:

$$V - 0.983\ VS - \frac{0.983\ VS}{2}$$

$$= V - 1.4745\ VS$$

$$= 3,723 - 1.4745 \times 3,273 \times 0.0521$$

$$= 3,437\ \text{Nm}^3/\text{min}$$

Fraction SO_2: 0.00096

$$(1 - 0.983)\ (3,723)\ (0.0521) \div 3,437$$

J.2.6 Air Pollution Impact

J.2.6.1 Table 7.4--SO_2 Control Alternative.

Baseline

Emissions/yr (total): 76,495 Mg/yr

MHR: 847 Mg/yr

$$\frac{71.2\text{ Mg S removed}}{\text{day}} \times \frac{350\text{ days}}{\text{year}} \times \frac{64\text{ Mg SO}_2}{32\text{ Mg S}}$$

$$\times \frac{0.017\text{ Mg emitted}}{\text{Mg to acid plant}} = 847$$

RV: $104.8 \times 350 \times \frac{64}{32} = 73{,}360$ Mg/yr

CV: $192.3 \times 350 \times \frac{64}{32} \times 0.017 = 2{,}288$

Control Alternative I-G

Emission/yr (total): 4,382 Mg/yr

MHR: 847 Mg/yr

CV: 2,288 Mg/yr

RV: $104.8 \times 350 \times \frac{64}{32} \times 0.017 = 1{,}247$

Emission reduction: $76{,}495 - 4{,}382 = 72{,}113$

Blister copper/yr: $1{,}364 \times 0.229 \times 350$ (99% recovery, 99% purity, 350 days/yr) $= 109{,}324$ Mg/yr

Reduction per unit of blister: $72{,}113 \times 1{,}000 \div 109{,}324 = 659$

J.2.6.2 Table 7-5--Fugitive Particulate Control.

Baseline, MHR: 568 Mg/yr

$$\frac{109{,}324\text{ Mg blister}}{\text{year}} \times \frac{5.2\text{ Kg fugitive}}{\text{Mg blister}} \times \frac{1\text{ Mg}}{1{,}000\text{ Kg}} = 568$$

Reduction, MHR: 506 Mg/yr

$$\frac{568\text{ Mg uncontrolled}}{\text{year}} \times \frac{0.90\text{ Mg captured}}{\text{Mg uncontrolled}}$$

$$\times \frac{0.99\text{ Mg collected}}{\text{Mg captured}} = 506$$

Controlled, MHR: $568 - 506 = 62$ Mg/yr

Control %: 506 ÷ 568 = 89 percent

Reduction: = 4.6 kg/Mg blister

$$\frac{506 \text{ Mg particulate}}{\text{Year}} \times \frac{1{,}000 \text{ Kg}}{\text{Mg}} \times \frac{\text{year}}{109{,}324 \text{ Mg blister}}$$

J.2.6.3 Table 7-7--Solid and Liquid Effluents from Gas Cleaning and Conditioning. Use factors from Appendix L.

Volume to acid plant: 3,725 Nm^3/min

or $3{,}725 \times 35.31 \times \frac{273}{294}$ = 122,100 scfm (°C)

$CaSO_4$: $122{,}100 \times 2.8 \times 10^{-5}$ = 3.4 Mg/yr

Liquid: $122{,}100 \times 1.8 \times 10^{-4}$ = 22.0 Mg/yr

J.2.6.4 Table 7-9--Solid and Liquid Wastes from FGDs. Use factors from Appendix L.

Volume to scrubber (Table 6-3): = 3,315 Nm^3/min at 12% SO_2

or

$3{,}315 \times 35.31 \times \frac{273}{294}$ = 108,700 scfm

$108{,}700 \times 1.7 \times 0.034$ = 6,282 Mg/yr solid waste

$108{,}700 \times 1.7 \times 1.8$ = 332,622 Mg/yr liquid waste

J.2.6.2.5 Table 7-12--energy impact. Energy requirements in Table 7-12 were estimated using relationships developed for the cost analysis (Chapter 8).

J.3 EXPANSION SCENARIOS

With the exception of converter analysis and distribution of acid plant flows to the existing single acid plant and a new double acid plant for scenarios requiring a new acid plant, the procedures for determining expansion scenario parameters are the same as those used for new greenfield smelters. Examples of each of these exceptions follow.

J.3.1 Converter Analysis

Using the same procedures used in Section J.2, the following converter profile, before dilution, is determined for Baseline II.

Converter aisle status

Converters on:

Slag blow	Copper blow	Offstack	Hr/day	Flow (Nm^3/min)	SO_2 (%)
2	1	0	4.0	1,550.8	12.3
2	0	1	5.2	1,142.2	11.1
1	1	1	9.5	979.6	13.0
1	0	2	2.3	553.6	11.1
0	1	2	1.7	408.5	15.8
0	0	3	1.3	0	-
Average				992.3	12.3

Dilution to attain 4.3 percent SO_2: $\frac{12.3}{4.3}$ = 2.860

Converter profile after dilution:

Converter aisle status

Converters on:

Slag blow	Copper blow	Off-stack	Hr/day	Nm^3/min	SO_2 (%)	O_2 (%)
2	1	0	4.0[a]	4,435	4.30	15.5
2	0	1	5.2	3,267	3.88	15.5
1	1	1	9.5	2,793	4.55	15.5
1	0	2	2.3	1,583	3.88	15.5
0	1	2	1.7	1,168	5.52	15.5
0	0	3	1.3			-
Average				2,838	4.3	15.5

---

[a] Example:

Flow: 1,550.8 × 2.860 = 4,435 Nm^3/min

SO_2: 12.3 ÷ 2.860 = 4.3 percent

O_2: Same procedure used in Section J.2.4

J.3.2 Acid Plant Flows

The procedures described herein apply to Expansion Scenarios 11 through 14. The example covers Scenario 11.

APPENDIX J

EXAMPLE CALCULATIONS
MODEL PLANT OPERATING PARAMETERS

Using the procedures described in Section J.2, the following parameters are developed:

	Flow (Nm^3/min)	SO_2 (%)	O_2 (%)
RV, to acid plant	1,440	3.0	15.6
Old CV	2,770	4.3	15.4
New CV	970	6.2	13.3

In this scenario, the flows from both the old and new converter are blended to smooth out variations encountered in individual converter operations. The new double acid plant is preferentially driven by sending a constant volume of the combined CV stream along with the RV stream to be controlled to the plant. The remainder of the combined CV stream is treated in the existing single acid plant.

Combined CV stream:

	Flow Nm^3/min	SO_2 (%)	O_2 (%)
Old CV	2,770	4.3	15.4
New CV	970	6.2	13.3
Combined	3,740	4.8	14.9

SO_2: $$\frac{2,770 \times 4.3 + 970 \times 6.2}{3,740} = 4.8$$

Equivalent new CV flow: $$\frac{970 \times 6.2}{4.8} = 1,253\ Nm^3/min$$

Double acid plant input:

	Flow Nm^3/min	SO_2 (%)	O_2 (%)
CV	1,253	4.8	14.9
RV	1,440	3.0	15.6
Combined	2,693	3.8	15.3

Single acid plant input: 3,740 - 1,253 = 2,487 Nm^3/min

APPENDIX K

MATHEMATICAL MODEL FOR ESTIMATING POSTEXPANSION REVERBERATORY GAS FLOW AND SO_2 CONCENTRATION FOR OXYGEN ENRICHMENT AND OXY-FUEL EXPANSION OPTIONS

APPENDIX K

MATHEMATICAL MODEL FOR ESTIMATING POST EXPANSION REVERBERATORY GAS FLOW AND SO_2 CONCENTRATION FOR OXYGEN ENRICHMENT AND OXY-FUEL EXPANSION OPTIONS

Assumptions:

1. Ten percent excess O_2.
2. Fifty percent of base case off-gases is dilution air. Amount of dilution air does not vary with expansion.
3. Fuel is equivalent to CH_4 for determining volume of combustion products.

Notation:

V = volume rate of off-gas

S = volume fraction of SO_2

C = volume of combustion products, excluding nitrogen, at theoretical O_2

O = volume rate of excess O_2

N = volume rate of nitrogen

P = volume fraction of oxygen in combustion air

E = Ratio of expansion capacity to base case capacity

H = ratio of expansion fuel to base case fuel

subscript b = base case

subscript e = expansion

Procedure:

1. V_b = dilution air + SO_2 + C_b + O_b + N_b

2. Dilution air $= \frac{V_b}{2}$

3. SO_2 rate $= V_bS_b$

4. $C_b + O_b + N_b = \frac{V_b}{2} - V_bS_b$ (from 1, 2, & 3)

$$= \frac{V_b(1-2S_b)}{2}$$

5. $CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$

Three volumes of combustion products require, at 10 percent excess O_2, 2.2 volumes of O_2 and $2.2 \times \frac{79}{21}$ or 8.3 volumes of nitrogen. This represents 0.2 volumes of excess O_2.

6. Using ratios of C_b , O_b, N_b determined in 4, the following relationships with $V_b + S_b$ are determined:

$$O_b = \frac{V_b\ (1-2S_b)}{2} \times \frac{0.2}{0.2 + 3 + 8.3} = 0.0087\ V_b\ (1-2S_b)$$

$$N_b = \frac{V_b\ (1-2S_b)}{2} \times \frac{8.3}{11.5} = 0.3609\ V_b\ (1-2S_b)$$

$$C_b = \frac{V_b\ (1-2S_b)}{2} \times \frac{3}{11.5} = 0.1304\ V_b\ (1-2S_b)$$

7. $V_e = \frac{V_b}{2} + SO_2 + C_e + O_e + N_e$

8. $SO_{2_e} = EV_bS_b$

9. $C_e = HC_b$

$= 0.1304\ HV_b\ (1-2S_b)$ (from 6)

10. $O_e = H\ O_b$

$$= 0.0087\ H\ V_b\ (1-2S_b) \qquad \text{(from 6)}$$

11. From 5, 3 volumes of combustion products require, at 10 percent excess O_2 (0.2 volumes), 2.2 volumes of oxygen and $2.2\ \frac{(1-P)}{P}$ volumes of nitrogen.

Each volume of combustion products is therefore associated with $2.2\ \frac{(1-P)}{3P}$ volumes of N_2.

$$N_2 = 0.1304\ HV_b\ (1-2S_b)\ 2.2\ \frac{(1-P)}{3P} \qquad \text{(from 9)}$$

$$= 0.0956\ HV_b\ (1-2S_b)\ \frac{(1-P)}{P}$$

12. Combining 7, 8, 9, 10, 11

$$V_e = \frac{V_b}{2} + EV_bS_b + 0.1304\ H\ V_b\ (1-2S_b) + 0.0087\ HV_b\ (1-2S_b)$$

$$+ 0.0956\ HV_b\ (1-2S_b)\ \frac{(1-P)}{P}$$

$$V_e = V_b \quad 0.5 + ES_b + H(1-2S_b)\ 0.1391 + 0.0956\ \frac{(1-P)}{P}$$

$$= V_b \left\{ 0.5 + ES_b + H(1-2S_b) \left[0.1391 + 0.0956\ \frac{(1-P)}{P} \right] \right\}$$

13. $$S_e = \frac{EV_b\ S_b}{V_e}$$

APPENDIX L

METHODOLOGY FOR ESTIMATING SOLID AND LIQUID WASTE DISPOSAL REQUIREMENTS*

L.1 GAS CLEANING AND CONDITIONING SYSTEMS

Scrubbing water purged from gas cleaning and conditioning equipment must be neutralized since this effluent is in fact a weak sulfuric acid solution. The effluent is neutralized via the following reaction:

$$H_2SO_4 + CaCO_3 \rightarrow CaSO_4 + CO_2 + H_2O.$$

Thus, as indicated, limestone ($CaCO_3$) is used as the neutralizing agent, producing calcium sulfate ($CaSO_4$), carbon dioxide gas, and water. Matthews et al.[1] report the following limestone usage rates, based upon 0.03 percent SO_3 in the inlet gas stream:

- 0.09 lb $CaCO_3/10^6$ scf (regenerative systems)
- 0.07 lb $CaCO_3/10^6$ scf (nonregenerative systems).

Thus, by noting the stoichiometry of the neutralization reaction, an expression that relates the amount of calcium sulfate produced and the volume of gas cleaned can be developed as follows:

Let V = the inlet gas stream volumetric flow rate in scfm (0° C)

$$CaSO_4 \text{ production rate} = V \times 0.09 \text{ (or } 0.07) \frac{\text{lb } CaCO_3}{10^6 \text{ scf}}$$

$$\times \frac{\text{lb} \cdot \text{mol } CaCO_3}{100 \text{ lb } CaCO_3} \times \frac{1 \text{ lb} \cdot \text{mol } CaSO_4 \text{ produced}}{\text{lb} \cdot \text{mol } CaCO_3 \text{ consumed}}$$

$$\times \frac{136 \text{ lbs } CaSO_4}{\text{lb} \cdot \text{mol } CaSO_4} \times \frac{60 \text{ min}}{\text{hr}} \times \frac{8{,}400 \text{ hrs of operation}}{\text{yr}}$$

---

*Because sources from which data were obtained used scfm at 0° C, the factors are calculated on this basis. Flows shown in this BID must be converted to this basis before using the factors.

$$\times \frac{1{,}000 \text{ g CaSO}_4}{2.2 \text{ lb CaSO}_4} \times \frac{\text{Mg}}{10^6 \text{ g}} \text{, in Mg per year}$$

$$\begin{aligned} &= 2.8 \times 10^{-5} \text{ V (regenerative), or} \\ &\quad 2.2 \times 10^{-5} \text{ V (nonregenerative).} \end{aligned}$$

Similar expressions can be derived for the water production rate:

$$\begin{aligned} \text{Water production rate} &= 3.7 \times 10^{-6} \text{ V (regenerative), or} \\ &\quad 2.9 \times 10^{-6} \text{ V (nonregenerative).} \end{aligned}$$

Once the $CaSO_4$ production rate is determined, stoichiometry can be envoked once again to determine the amount of acid neutralized:

$$2.8 \times 10^{-5} \text{ V } \frac{\text{Mg CaSO}_4}{\text{yr}} \times \frac{10^6 \text{ g}}{\text{Mg}} \times \frac{\text{g} \cdot \text{mol H}_2\text{SO}_4 \text{ neutralized}}{1 \text{ g} \cdot \text{mol CaSO}_4 \text{ formed}}$$

$$\times \frac{98 \text{ g H}_2\text{SO}_4}{\text{g} \cdot \text{mol}} \times \frac{\text{Mg}}{10^6 \text{ g}} \times \frac{\text{g} \cdot \text{mole CaSO}_4}{136 \text{ g CaSO}_4}\text{, in Mg per year}$$

$$\begin{aligned} &= 2.0 \times 10^{-5} \text{ V (regenerative), or} \\ &\quad 1.6 \times 10^{-5} \text{ V (nonregenerative).} \end{aligned}$$

Then, once the amount of acid neutralized is determined, the total amount of liquid (calculated as water) requiring disposal can be estimated as follows:

Noting that the purge is normally about 10 percent H_2SO_4 by weight,

$$\text{Total liquid effluent rate} = 2.0 \times 10^{-5} \text{ V} \times \frac{(1-0.10)}{0.10}$$

$$+ 3.7 \times 10^{-6} \text{ V} = (18 \times 10^{-5} + 3.7 \times 10^{-6}) \text{ V, in Mg per year}$$

$$\begin{aligned} &= 1.8 \times 10^{-4} \text{ V (regenerative), or} \\ &\quad 1.5 \times 10^{-4} \text{ V (nonregenerative).} \end{aligned}$$

L.2 LIME/LIMESTONE SLURRY SCRUBBING PROCESS

Matthews et al.[1] report that sludge consisting primarily of calcium sulfite ($CaSO_3$) is produced at the rate of 6 to 7 lbs per lb of SO_2 absorbed. Thus, the rate of sludge generation can be estimated as follows:

$$\text{Sludge generation rate} = V\ (\text{scfm}) \times \frac{C\ (\%\ SO_2)}{100} \times \frac{\text{lb} \cdot \text{mol}\ SO_2}{359\ \text{ft}^3\ SO_2}$$

$$\times \frac{64\ \text{lb}\ SO_2}{\text{lb} \cdot \text{mol}\ SO_2} \times \frac{\eta}{100} \times \frac{60\ \text{min}}{\text{hr}} \times \frac{8{,}400\ \text{hr}}{\text{yr}} \times \frac{6\ \text{lb sludge}}{\text{lb} \cdot SO_2\ \text{absorbed}}$$

$$\times \frac{1{,}000\ \text{g}}{2.2\ \text{lb}} \times \frac{\text{Mg}}{10^6\ \text{g}},$$ in Mg per year, where η = the FGD SO_2 removal efficiency (90 percent).

Condensing terms yields a sludge generation rate (Mg/yr) of 2.2 VC.

Typically, a mixture of 15 weight percent sludge[1] and 85 weight percent water is ponded; therefore, the amount of liquid that must be pumped to the pond can be estimated as follows:

$$\text{Liquid effluent rate} = 2.2\ VC \times \frac{(1-0.15)}{0.15}$$

$$= 13\ VC,\ \text{in Mg/yr}.$$

L.3 MAGOX SLURRY SCRUBBING PROCESS

A scrubber costs program developed by PEDCo[2] was used to estimate the amount of absorbent purge taken from the MAGOX system. This was done by assuming that the amount of solid material purged (assumed to be $MgSO_3$) is stoichiometrically equivalent to the amount of MgO fed as makeup. The subroutine SMTMG1 from the PEDCo program calculates the MgO makeup feed rate as follows:

a. SO_2 feed rate (lb/hr) = $V\ (\text{scfm}) \times \frac{C(\%\ SO_2)}{100} \times \frac{60 \times 64}{386.7}$

b. SO_2 absorbed (lb/hr) = a. $\times \frac{SO_2\ \text{removal efficiency}}{100}$

c. MgO actually consumed (lb/hr) = b. × 0.625

d. MgO required (lb/hr) = c. $\times 1.1 \times \frac{100}{SO_2\ \text{removal efficiency}}$

e. Makeup MgO (lb/hr) = d. × 0.05

As indicated by expression "e" above, the MgO makeup rate is expressed as 5 percent of the MgO actually required. Assuming an SO_2 removal efficiency of 90 percent, Equations (a) through (e) can be condensed to yield the following expression:

$$\text{MgO makeup rate} = 0.0034 \text{ VC, in lb MgO per hr.}$$

Converting to Mg per year yields a MgO makeup rate (Mg/yr) = 0.013 VC. As mentioned previously, it has been assumed that the amount of solid material purged (assumed to be $MgSO_3$) is stoichiometrically equivalent to the amount of MgO fed as makeup. The following reaction was chosen to be representative of the stoichiometry involved:

$$MgO + SO_2 + 3H_2O \rightarrow MgSO_3 \cdot 6H_2O.$$

As indicated, the $MgSO_3$ is in the form of a hydrated crystal. Since one mol of MgO is consumed in the formation of one mole of $MgSO_3 \cdot 6H_2O$, an expression for the $MgSO_3 \cdot 6H_2O$ purge rate can be developed as follows:

$$0.0034 \text{ VC } \frac{\text{lb MgO}}{\text{hr}} \times \frac{\text{lb} \cdot \text{mol MgO}}{40 \text{ lb MgO}} \times \frac{1 \text{ lb} \cdot \text{mol } MgSO_3}{\text{lb} \cdot \text{mol MgO}}$$

$$\times \frac{104 \text{ lb } MgSO_3}{\text{lb} \cdot \text{mol } MgSO_3} \times \frac{1{,}000 \text{ g}}{2.2 \text{ lb } MgSO_3} \times \frac{\text{Mg}}{10^6 \text{ g}}$$

$$\times \frac{8{,}400 \text{ hr}}{\text{yr}} = 0.034 \text{ VC , in Mg per year.}$$

Next, assuming that the mixture to be ponded is 2 weight percent solids,[3] an expression for the liquid effluent rate can be developed as follows:

$$\text{Liquid effluent rate (Mg/yr)} = 0.034 \text{ VC} \times \frac{0.02}{1-0.02}$$

$$= 1.7 \text{ VC.}$$

L.4 PARTICULATE MATTER CONTROL ON REVERBERATORY SMELTING FURNACES

To assess the impact of the evaporative cooling procedure on the gas stream volumetric flow rate, an energy balance is used to estimate the amount of water that is evaporated in reducing the gas stream temperature from 400° C to 100° C. The energy balance has the following form:

$$- \dot{m}_G \hat{C}_{P_G} \Delta T_G = \dot{m}_W \Delta \hat{H}_{V_W}, \qquad (L-1)$$

where

$\dot{m}_G$ = the molar flow rate of the gas stream

$\hat{C}_{P_G}$ = the specific heat capacity of the gas stream

ΔT_G = the temperature change associated with cooling the gas stream

$\dot{m}_W$ = the mols of water evaporated per unit time

$\Delta \hat{H}_{V_W}$ = the latent heat of vaporization of water at 100° C.

Since $\dot{m}_W$ is the quantity of interest, Eq. (L-1) can be rearranged to yield:

$$\dot{m}_W = \frac{- \dot{m}_G \hat{C}_{P_G} \Delta T_G}{\Delta \hat{H}_{V_W}} . \qquad (L-2)$$

An average specific heat capacity for the reverberatory furnace offgas stream can be estimated using the following gas-stream composition:

Component	Vol %
SO_2	1.0
O_2	11.0
N_2	
H_2O	
CO_2	

The average heat capacity can then be calculated as follows:

$$\hat{C}_{P_G} = \Sigma\, y_i\, \hat{C}_{P_i}\ ,$$

where

y_i = the gas stream volume fraction of species i

$\hat{C}_{P_i}$ = the specific heat capacity of species i.

The following pure component specific heat capacities are used:

Component	Specific heat capacity @ 25° C
SO_2	39.8 J/g · mol °K
O_2	29.4 J/g · mol °K
N_2	29.1 J/g · mol °K
H_2O	33.6 J/g · mol °K
CO_2	37.1 J/g · mol °K

$$\hat{C}_{P_G} \cong (0.012)\ (39.8) + (0.099)\ (29.4)$$

$$+ (0.083)\ (29.1) + (0.096)\ (33.6)$$

$$+ (0.044)\ (37.1) = 32.4 \text{ J/g} \cdot \text{mol °K}.$$

Also,

$$\Delta T_G = -300° \text{ C}$$

$$\Delta\hat{H}_{V_W} = 40.7 \text{ kJ/g} \cdot \text{mol @ } 100° \text{ C}$$

Now, $\dot{m}_W$ can be estimated using Equation (L-2).

L.5 REFERENCES

1. Matthews, J. C., F. L. Bellegia, C. H. Gooding, and G. E. Weant. SO_2 Control Processes for Nonferrous Smelters. Research Triangle Institute, Research Triangle Park, N.C. Publication No. EPA-600/2-76-008. January 1976.

2. PEDCo Environmental, Inc., "Users Guide, Computerized Approach to Estimating SO_2 Scrubber Costs at Nonferrous Smelters, "EPA Contract No. 68-03-2924, April 1982.

3. Anderson, K. P., et al., "Definitive SO_x Control Process Evaluations: Limestone, Lime, and Magensia FGD Processes," TVA ECDP B-7, January 1980.

APPENDIX L

METHODOLOGY FOR ESTIMATING SOLID AND LIQUID WASTE DISPOSAL REQUIREMENTS

APPENDIX M

DETAILED COSTS FOR GREENFIELD SMELTERS

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : Base Case
Plant Scenario : New

Date : 09/14/82
Time : 12:17

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	162.000,000.	46.278.400.
Annualized Costs		
Raw materials	688.625.	19.922.
Process water	105.555.	134.899.
Cooling water	0.	93.526.
Electricity	349.860.	6.471,850.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	9.538.830.	0.
Solids disposal	0.	0.
Labor: Direct Operating	1.538.450.	309.812.
Supervision	307.690.	61.962.
Maint.: Labor & Matl.	6.480.000.	1.851.140.
Supervision	972.000.	277.670.
Overhead	4.649.070.	1,250.290.
Taxes. ins.. admin.	6.480.000.	1.851.140.
	----------	---------
Total Operating Cost	31.110.100.	12.322.200.
Capital Recovery Cost	26.365.500.	7.531.810.
	----------	---------
Annualized Cost	57.475.600.	19.854.000.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : 45% Blending + DC/DA (I-A)
Plant Scenario : New

Date : 02/23/83
Time : 10:18

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	0.	61,187,100.
Annualized Costs		
Raw materials	0.	28,660.
Process water	0.	164,999.
Cooling water	0.	134,550.
Electricity	0.	9,310,590.
Supp. heat (Nat. gas)	0.	6,700.
Bunker C Fuel Oil	0.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	309,812.
Supervision	0.	61,962.
Maint.: Labor & Matl.	0.	2,447,480.
Supervision	0.	367,122.
Overhead	0.	1,593,190.
Taxes, ins., admin.	0.	2,447,480.
Total Operating Cost	0.	16,872,500.
Capital Recovery Cost	0.	9,958,190.
Annualized Cost	0.	26,830,700.

Negative values indicate savings over base case costs.

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : MgO FGD + DC/DA (I-B)
Plant Scenario : New

Date : 09/14/82
Time : 12:18

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

```
                                   Process       Control
Capital Cost                            0.   73.996.900.


Annualized Costs

Raw materials                           0.      598.641.
Process water                           0.      220.938.
Cooling water                           0.      107.538.
Electricity                             0.    7.221,550.
Supp. heat (Nat. gas)                   0.            0.
Bunker C Fuel Oil                       0.    3.733.570.
Solids disposal                         0.            0.
Labor: Direct Operating                 0.      557.662.
       Supervision                      0.      111.532.
Maint.: Labor & Matl.                   0.    2.959.880.
        Supervision                     0.      443.982.
Overhead                                0.    2.036.530.
Taxes. ins., admin.                     0.    2.959.880.
                                ----------    ---------
Total Operating Cost                    0.   20.951.700.

Capital Recovery Cost                   0.   12.043.000.
                                ----------    ---------
Annualized Cost                         0.   32.994.700.
```

Negative values indicate savings over base case costs.

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : NH3 FGD + DC/DA (I-C)
Plant Scenario : New

Date : 09/14/82
Time : 12:18

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	0.	62.839.200.
Annualized Costs		
Raw materials	0.	4.664.440.
Process water	0.	147.141.
Cooling water	0.	544.955.
Electricity	0.	8.799.440.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	0.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	743.549.
Supervision	0.	148.710.
Maint.: Labor & Matl.	0.	2.513.570.
Supervision	0.	377.035.
Overhead	0.	1.891.430.
Taxes, ins., admin.	0.	2.513,570.
Total Operating Cost	0.	22.343,800.
Capital Recovery Cost	0.	10.227.100.
Annualized Cost	0.	32.570.900.

Negative values indicate savings over base case costs.

Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : LL FGD + DC/DA (I-D)
Plant Scenario : New

Date : 09/14/82
Time : 12:19

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	0.	69.742.000.
Annualized Costs		
Raw materials	0.	1.243.260.
Process water	0.	217.313.
Cooling water	0.	93.526.
Electricity	0.	6.752.510.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	0.	0.
Solids disposal	0.	820.821.
Labor: Direct Operating	0.	557.662.
Supervision	0.	111.532.
Maint.: Labor & Matl.	0.	2.789,680.
Supervision	0.	418.452.
Overhead	0.	1.938.660.
Taxes, ins., admin.	0.	2.789.680.
	----------	---------
Total Operating Cost	0.	17.733,100.
Capital Recovery Cost	0.	11.350.500.
	----------	---------
Annualized Cost	0.	29.083,600.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : 100% Blending + DC/DA (I-E)
Plant Scenario : New

Date : 02/02/83
Time : 13:31

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	0.	74,231,100.
Annualized Costs		
Raw materials	0.	39,322.
Process water	0.	226,381.
Cooling water	0.	184,605.
Electricity	0.	11,527,700.
Supp. heat (Nat. gas)	0.	72,600.
Bunker C Fuel Oil	0.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	0.	2,969,240.
Supervision	0.	445,387.
Overhead	0.	1,874,610.
Taxes, ins., admin.	0.	2,969,240.
Total Operating Cost	0.	20,643,700.
Capital Recovery Cost	0.	12,081,100.
Annualized Cost	0.	32,724,800.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : Oxygen enrichment + DC/DA (I-F)
Plant Scenario : New

Date : 02/07/83
Time : 09:49

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	0.	67,215,300.
Annualized Costs		
Raw materials	0.	1,941,860.
Process water	0.	193,775.
Cooling water	0.	158,016.
Electricity	0.	9,867,310.
Supp. heat (Nat. gas)	0.	5,700.
Bunker C Fuel Oil	-1,716,990.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	0.	2,688,610.
Supervision	0.	403,292.
Overhead	0.	1,713,250.
Taxes, ins., admin.	0.	2,688,610.
Total Operating Cost	-1,716,990.	19,995,000.
Capital Recovery Cost	0.	10,939,300.
Annualized Cost	-1,716,990.	30,934,300.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Not Applicable
Control Option : Oxy-fuel burners + DC/DA (I-G)
Plant Scenario : New

Date : 10/15/82
Time : 14:10

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	0.	55.216.600.
Annualized Costs		
Raw materials	0.	3.630.020.
Process water	0.	186.034.
Cooling water	0.	128.978.
Electricity	0.	8.054.090.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-3.815.530.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278.831.
Supervision	0.	55.766.
Maint.: Labor & Matl.	0.	2.208.660.
Supervision	0.	331.299.
Overhead	0.	1.437.280.
Taxes, ins., admin.	0.	2.208.660.
Total Operating Cost	-3.815.530.	18.519.600.
Capital Recovery Cost	0.	8.986.490.
Annualized Cost	-3.815,530.	27.506.100.

Negative values indicate savings over base case costs.

Fugitive Control Costs

Annualized Costs ($ 1000's, June 1981)

Plant	Capture System		Collection System		C[illegible] Collection	
	w/BE	w/AC	w/BE	w/AC	w/BE	w/AC
Greenfield Plants						
MHR-RV-CV						
MHR	39	39	234	234	273	273
RV	103	103	533	533	636	636
CV	1,713	2,237	4,185	1,401	5,898	3,638
FF-CV						
FF	78	78	358	358	436	436
CV	1,713	2,237	4,185	1,401	5,898	3,638
Expansion Base Cases						
I MHR-RV-CV						
MHR	39	39	234	234	273	273
RV	103	103	533	533	636	636
CV	1,713	2,237	4,185	1,401	5,898	3,638
II RV-CV						
RV	103	103	533	533	636	636
CV	1,713	2,237	4,185	1,401	5,898	3,638
III FBR-RV-CV						
FBR	0	0	0	0	0	0
RV	103	103	533	533	636	636
CV	1,713	2,237	4,185	1,401	5,898	3,638
IV EF-CV						
EF	103	103	491	491	594	594
CV	1,713	2,237	4,185	1,401	5,898	3,638
V FF-CV						
FF	78	78	358	358	436	436
ESCF	66	66	491	491	557	557
CV	1,713	2,237	4,185	1,401	5,898	3,638
Expansion Options						
9-13						
CV	0	746	42	579	42	1,325
18						
FBR	0	0	0	0	0	0
CV	0	746	42	579	42	1,325

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- All labor costs are assigned to the collection system (baghouse)
- Electrical usage rate is calculated as $2.5 * 10^4$ kwh/yr-1000 scfm

Fugitive Control Costs

Capital Costs ($ 1000's, June 1981)

Plant	Flow (1000's SCFM)		Capture System			Collection System (Baghouse)	
			Hoods and ducting		Air Curtain		
	w/BE	w/AC	w/BE	w/AC		w/BE	w/AC
Greenfield Plants							
MHR-RV-CV							
MHR	20	20	116	116	0	571	571
RV	65	65	298	298	0	1,539	1.539
CV	750	200	5.300	1,723	6,170	12,213	4,133
FF-CV							
FF	45	45	224	224	0	1,130	1.130
CV	750	200	5,300	1,723	6.170	12.213	4.133
Expansion Base Cases							
I MHR-RV-CV							
MHR	20	20	116	116	0	571	571
RV	65	65	298	298	0	1.539	1,539
CV	750	200	5.300	1,723	6,170	12,213	4.133
II RV-CV							
RV	65	65	298	298	0	1.539	1.539
CV	750	200	5.300	1,723	6.170	12.213	4.133
III FBR-RV-CV							
FBR	0	0	0	0	0	0	0
RV	65	65	298	298	0	1,539	1,539
CV	750	200	5.300	1,723	6.170	12,213	4,133
IV EF-CV							
EF	65	65	298	298	0	1,539	1.539
CV	750	200	5.300	1,723	6.170	12,213	4.133
V FF-CV							
FF	45	45	224	224	0	1.130	1.130
ESCF	65	65	239	239	0	1,539	1.539
CV	750	200	5.300	1,723	6.170	12.213	4.133
Expansion Options							
9-13							
CV	0	67	0	575	2.057	0	1,647
13							
FBR	0	0	0	0	0	0	0
CV	0	67	0	575	2.057	0	1.647

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[1] No fugitive controls are required on a fluid bed roaster. (See Section 4.7.4)

[2] It is assumed that a new converter would be added in an existing building. Since the building evacuation cost is a function only of building size, no new cost would be incurred for fugitive control with Building Evacuation.

[3] This is 1/3 of the ASARCO-Tacoma air curtain design flow rate.

APPENDIX N

FUGITIVE EMISSION CONTROL COSTS

APPENDIX O

DETAILED COSTS FOR EXPANSION SCENARIOS

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Oxygen enrichment
Control Option : PB - SC/SA
Plant Scenario : 1

Date : 02/04/83
Time : 01:06

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	10,907,600.
Annualized Costs		
Raw materials	2,635,430.	9,529.
Process water	0.	17,322.
Cooling water	0.	44,737.
Electricity	9,534.	1,641,510.
Supp. heat (Nat. gas)	0.	60,000.
Bunker C Fuel Oil	341,201.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	0.
Supervision	0.	0.
Maint.: Labor & Matl.	20,400.	436,303.
Supervision	3,060.	65,445.
Overhead	11,730.	250,874.
Taxes, ins., admin.	20,400.	436,303.
Total Operating Cost	3,041,750.	2,962,020.
Capital Recovery Cost	83,002.	1,775,210.
Annualized Cost	3,124,750.	4,737,230.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Oxygen enrichment
Control Option : LL - SC/SA
Plant Scenario : 2

Date : 02/04/83
Time : 00:01

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	15,559,600.
Annualized Costs		
Raw materials	2,635,430.	230,270.
Process water	0.	24,504.
Cooling water	0.	20,258.
Electricity	9,534.	1,041,000.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	341,201.	0.
Solids disposal	0.	151,609.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	622,384.
Supervision	3,060.	93,358.
Overhead	11,730.	525,170.
Taxes, ins., admin.	20,400.	622,384.
Total Operating Cost	3,041,750.	3,665,530.
Capital Recovery Cost	83,002.	2,532,330.
Annualized Cost	3,124,750.	6,197,860.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	MHR-RV-CV	**Date :** 02/04/83
Expansion Option :	Oxygen enrichment	**Time :** 00:02
Control Option :	MgO - SC/SA	
Plant Scenario :	3	

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	17,589,600.
Annualized Costs		
Raw materials	2,635,430.	118,932.
Process water	0.	23,371.
Cooling water	0.	23,635.
Electricity	9,534.	1,056,680.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	341,201.	725,197.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	703,583.
Supervision	3,060.	105,537.
Overhead	11,730.	571,858.
Taxes, ins., admin.	20,400.	703,583.
Total Operating Cost	3,041,750.	4,366,970.
Capital Recovery Cost	83,002.	2,862,700.
Annualized Cost	3,124,750.	7,229,670.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : MHR-RV-CV
Expansion Option : Oxygen enrichment
Control Option : NH3 - SC/SA
Plant Scenario : 4

Date : 02/08/83
Time : 09:01

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	13,462,000.
Annualized Costs		
Raw materials	2,635,430.	909,010.
Process water	0.	11,402.
Cooling water	0.	110,237.
Electricity	9,534.	1,428,140.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	341,201.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	464,718.
Supervision	0.	92,944.
Maint.: Labor & Matl.	20,400.	538,479.
Supervision	3,060.	80,772.
Overhead	11,730.	558,456.
Taxes, ins., admin.	20,400.	538,479.
Total Operating Cost	3,041,750.	4,762,640.
Capital Recovery Cost	83,002.	2,190,940.
Annualized Cost	3,124,750.	6,953,580.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : RV-CV
Expansion Option : Oxygen enrichment
Control Option : PB - SC/SA
Plant Scenario : 7

Date : 02/04/83
Time : 16:48

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	11,360,900.
Annualized Costs		
Raw materials	2,690,980.	10,608.
Process water	0.	12,473.
Cooling water	0.	49,802.
Electricity	4,662.	1,631,450.
Supp. heat (Nat. gas)	0.	14,300.
Bunker C Fuel Oil	-158,760.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	0.
Supervision	0.	0.
Maint.: Labor & Matl.	20,400.	454,436.
Supervision	3,060.	68,165.
Overhead	11,730.	261,301.
Taxes, ins., admin.	20,400.	454,436.
Total Operating Cost	2,592,470.	2,956,980.
Capital Recovery Cost	83,002.	1,848,990.
Annualized Cost	2,675,470.	4,805,960.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : RV-CV
Expansion Option : Oxygen enrichment
Control Option : LL - SC/SA
Plant Scenario : 8

Date :02/04/83
Time : 15:46

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	17,507,900.
Annualized Costs		
Raw materials	2,690,980.	414,896.
Process water	0.	35,402.
Cooling water	0.	25,323.
Electricity	4,662.	1,135,700.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-158,760.	0.
Solids disposal	0.	274,763.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	700,317.
Supervision	3,060.	105,047.
Overhead	11,730.	569,981.
Taxes, ins., admin.	20,400.	700,317.
Total Operating Cost	2,592,470.	4,296,340.
Capital Recovery Cost	83,002.	2,849,410.
Annualized Cost	2,675,470.	7,145,760.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	RV-CV	Date : 02/04/83	
Expansion Option :	Oxygen enrichment	Time : 16:23	
Control Option :	MgO - SC/SA		
Plant Scenario :	9		

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	20,204,900.
Annualized Costs		
Raw materials	2,690,980.	207,963.
Process water	0.	31,984.
Cooling water	0.	29,544.
Electricity	4,662.	1,164,310.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-158,760.	1,314,500.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	808,195.
Supervision	3,060.	121,229.
Overhead	11,730.	632,011.
Taxes, ins., admin.	20,400.	808,195.
Total Operating Cost	2,592,470.	5,452,520.
Capital Recovery Cost	83,002.	3,288,350.
Annualized Cost	2,675,470.	8,740,870.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : RV-CV
Expansion Option : Oxygen enrichment
Control Option : NH3 - SC/SA
Plant Scenario : 10

Date : 02/08/83
Time : 09:03

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	15,576,000.
Annualized Costs		
Raw materials	2,690,980.	1,639,520.
Process water	0.	9,703.
Cooling water	0.	183,986.
Electricity	4,662.	1,842,400.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-158,760.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	464,718.
Supervision	0.	92,944.
Maint.: Labor & Matl.	20,400.	623,040.
Supervision	3,060.	93,456.
Overhead	11,730.	637,079.
Taxes, ins., admin.	20,400.	623,040.
Total Operating Cost	2,592,470.	6,209,890.
Capital Recovery Cost	83,002.	2,534,990.
Annualized Cost	2,675,470.	8,744,880.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	RV-CV	Date : 02/04/83
Expansion Option :	Oxy-fuel burners	Time : 15:47
Control Option :	PB - DC/DA	
Plant Scenario :	11	

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	32,800,000.	40,013,200.
Annualized Costs		
Raw materials	7,031,760.	16,002.
Process water	0.	92,126.
Cooling water	0.	75,125.
Electricity	11,634.	4,691,190.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-992,250.	0.
Solids disposal	0.	0.
Labor: Direct Operating	92,944.	278,831.
Supervision	18,589.	55,766.
Maint.: Labor & Matl.	1,312,000.	1,600,530.
Supervision	196,800.	240,079.
Overhead	810,166.	1,087,600.
Taxes, ins., admin.	1,312,000.	1,600,530.
Total Operating Cost	9,793,650.	9,737,780.
Capital Recovery Cost	5,338,200.	6,512,150.
Annualized Cost	15,131,800.	16,249,900.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	RV-CV	**Date :** 02/04/83	
Expansion Option :	Oxy-fuel burners	**Time :** 15:48	
Control Option :	LL - DC/DA		
Plant Scenario :	12		

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	32,800,000.	41,321,200.
Annualized Costs		
Raw materials	7,031,760.	1,021,190.
Process water	0.	110,739.
Cooling water	0.	34,608.
Electricity	11,634.	2,579,510.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-992,250.	0.
Solids disposal	0.	680,238.
Labor: Direct Operating	92,944.	557,662.
Supervision	18,589.	111,532.
Maint.: Labor & Matl.	1,312,000.	1,652,850.
Supervision	196,800.	247,927.
Overhead	810,166.	1,284,980.
Taxes, ins., admin.	1,312,000.	1,652,850.
	----------	---------
Total Operating Cost	9,793,650.	9,944,080.
Capital Recovery Cost	5,338,200.	6,725,020.
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Annualized Cost	15,131,800.	16,669,100.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	RV-CV	Date : 02/04/83
Expansion Option :	Oxy-fuel burners	Time : 15:48
Control Option :	MgO - DC/DA	
Plant Scenario :	13	

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	32,800,000.	47,461,100.
Annualized Costs		
Raw materials	7,031,760.	494,012.
Process water	0.	128,387.
Cooling water	0.	47,270.
Electricity	11,634.	3,196,520.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-992,250.	3,185,120.
Solids disposal	0.	0.
Labor: Direct Operating	92,944.	557,662.
Supervision	18,589.	111,532.
Maint.: Labor & Matl.	1,312,000.	1,898,440.
Supervision	196,800.	284,767.
Overhead	810,166.	1,426,200.
Taxes, ins., admin.	1,312,000.	1,898,440.
Total Operating Cost	9,793,650.	13,228,400.
Capital Recovery Cost	5,338,200.	7,724,290.
Annualized Cost	15,131,800.	20,952,600.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : RV-CV
Expansion Option : Oxy-fuel burners
Control Option : NH3 - DC/DA
Plant Scenario : 14

Date : 02/08/83
Time : 09:04

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	32,800,000.	37,415,000.
Annualized Costs		
Raw materials	7,031,760.	3,962,480.
Process water	0.	69,228.
Cooling water	0.	420,035.
Electricity	11,634.	4,648,410.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-992,250.	0.
Solids disposal	0.	0.
Labor: Direct Operating	92,944.	743,549.
Supervision	18,589.	148,710.
Maint.: Labor & Matl.	1,312,000.	1,496,600.
Supervision	196,800.	224,490.
Overhead	810,166.	1,306,680.
Taxes, ins., admin.	1,312,000.	1,496,600.
Total Operating Cost	9,793,650.	14,516,800.
Capital Recovery Cost	5,338,200.	6,089,300.
Annualized Cost	15,131,800.	20,606,100.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	RV-CV		Date : 02/02/83
Expansion Option :	Calcine charge		Time : 09:02
Control Option :	DC/DA		
Plant Scenario :	15		

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	44,000,000.	26,841,300.
Annualized Costs		
Raw materials	467,850.	10,428.
Process water	0.	81,002.
Cooling water	0.	48,958.
Electricity	595,560.	3,057,180.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	3,828,080.	0.
Solids disposal	0.	0.
Labor: Direct Operating	464,718.	278,831.
Supervision	92,944.	55,766.
Maint.: Labor & Matl.	1,760,000.	1,073,650.
Supervision	264,000.	161,048.
Overhead	1,290,830.	784,648.
Taxes, ins., admin.	1,760,000.	1,073,650.
Total Operating Cost	10,524,000.	6,625,160.
Capital Recovery Cost	7,161,000.	4,368,420.
Annualized Cost	17,685,000.	10,993,600.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : FBR-RV-CV
Expansion Option : Oxygen enrichment
Control Option : PB - SC/SA
Plant Scenario : 18

Date :02/04/83
Time : 15:49

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	2,938,390.
Annualized Costs		
Raw materials	2,000,080.	5,574.
Process water	0.	31,376.
Cooling water	0.	26,167.
Electricity	89,040.	1,049,770.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-1,525,680.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	0.
Supervision	0.	0.
Maint.: Labor & Matl.	20,400.	117,536.
Supervision	3,060.	17,630.
Overhead	11,730.	67,583.
Taxes, ins., admin.	20,400.	117,536.
Total Operating Cost	619,021.	1,433,170.
Capital Recovery Cost	83,002.	478,223.
Annualized Cost	702,024.	1,911,390.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : FBR-RV-CV
Expansion Option : Oxygen enrichment
Control Option : LL - SC/SA
Plant Scenario : 19

Date : 02/04/83
Time : 15:49

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	5,700,670.
Annualized Costs		
Raw materials	2,000,080.	138,555.
Process water	0.	41,495.
Cooling water	0.	5,909.
Electricity	89,040.	463,602.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-1,525,680.	0.
Solids disposal	0.	92,121.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	228,027.
Supervision	3,060.	34,204.
Overhead	11,730.	298,414.
Taxes, ins., admin.	20,400.	228,027.
Total Operating Cost	619,021.	1,864,950.
Capital Recovery Cost	83,002.	927,783.
Annualized Cost	702,024.	2,792,730.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : FBR-RV-CV
Expansion Option : Oxygen enrichment
Control Option : MgO - SC/SA
Plant Scenario : 20

Date : 02/04/83
Time : 15:50

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	6,970,250.
Annualized Costs		
Raw materials	2,000,080.	72,145.
Process water	0.	40,177.
Cooling water	0.	6,753.
Electricity	89,040.	408,034.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-1,525,680.	441,670.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	278,810.
Supervision	3,060.	41,822.
Overhead	11,730.	327,614.
Taxes, ins., admin.	20,400.	278,810.
Total Operating Cost	619,021.	2,230,430.
Capital Recovery Cost	83,002.	1,134,410.
Annualized Cost	702,024.	3,364,840.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type :	FBR-RV-CV		Date : 02/08/83
Expansion Option :	Oxygen enrichment		Time : 09:05
Control Option :	NH3 - SC/SA		
Plant Scenario :	21		

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	2,906,600.
Annualized Costs		
Raw materials	2,000,080.	553,150.
Process water	0.	32,743.
Cooling water	0.	58,652.
Electricity	89.040.	575,345.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-1,525,680.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	464,718.
Supervision	0.	92,944.
Maint.: Labor & Matl.	20,400.	116,267.
Supervision	3,060.	17,440.
Overhead	11,730.	345,685.
Taxes, ins., admin.	20,400.	116,267.
Total Operating Cost	619,021.	2,373,210.
Capital Recovery Cost	83,002.	473,063.
Annualized Cost	702,024.	2,846,270.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : EF-CV
Expansion Option : Calcine charge
Control Option : DC/DA
Plant Scenario : 23

Date : 02/02/83
Time : 09:04

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	46,600,000.	37,504,900.
Annualized Costs		
Raw materials	467,850.	14,564.
Process water	0.	83,845.
Cooling water	0.	68,372.
Electricity	583,800.	4,269,510.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	3,666,090.	0.
Solids disposal	0.	0.
Labor: Direct Operating	371,774.	278,831.
Supervision	74,355.	55,766.
Maint.: Labor & Matl.	1,864,000.	1,500,190.
Supervision	279,600.	225,029.
Overhead	1,294,860.	1,029,910.
Taxes, ins., admin.	1,864,000.	1,500,190.
Total Operating Cost	10,466,300.	9,026,220.
Capital Recovery Cost	7,584,150.	6,103,920.
Annualized Cost	18,050,500.	15,130,100.

Negative values indicate savings over base case costs.

Copper Smelter Costs

Plant type : FF-CV
Expansion Option : Oxygen enrichment
Control Option : DC/DA
Plant Scenario : 26

Date : 02/04/83
Time : 16:20

Process costs include new hardware associated with copper production. For the greenfield smelter, the process cost is the Baseline Case Cost (Smelter plus fugitive capture). For the expansion scenarios, process costs include any new roaster or converter. Control costs include all equipment associated with emission reduction. Oxygen enrichment and oxyfuel costs are considered as control costs along with acid plant and FGD costs for the greenfield smelter. Oxygen enrichment and oxyfuel costs are considered as expansion costs for existing plants.

	Process	Control
Capital Cost	510,000.	5,088,630.
Annualized Costs		
Raw materials	1,457,230.	3,956.
Process water	0.	30,725.
Cooling water	0.	18,570.
Electricity	5,817.	1,159,620.
Supp. heat (Nat. gas)	0.	0.
Bunker C Fuel Oil	-158,760.	0.
Solids disposal	0.	0.
Labor: Direct Operating	0.	278,831.
Supervision	0.	55,766.
Maint.: Labor & Matl.	20,400.	203,545.
Supervision	3,060.	30,532.
Overhead	11,730.	284,337.
Taxes, ins., admin.	21,930.	218,811.
Total Operating Cost	1,361,400.	2,284,690.
Capital Recovery Cost	83,002.	828,175.
Annualized Cost	1,444,400.	3,112,870.

Negative values indicate savings over base case costs.

APPENDIX P

METHODOLOGY UTILIZED TO DETERMINE THE COSTS ASSOCIATED WITH SULFURIC ACID PLANT PREHEATER OPERATION

APPENDIX P

METHODOLOGY UTILIZED TO DETERMINE THE COSTS ASSOCIATED WITH SULFURIC ACID PLANT PREHEATER OPERATION

P.1 DETERMINATION OF THE STANDARD HEAT OF REACTION

(@ 298 K) for the conversion reaction.

$$SO_2 + 1/2\ O_2 \rightarrow SO_3$$

ΔH^o_R = the standard heat of reaction at 1 atmosphere and 298 K

$$= \sum_P \nu_i \, \Delta H^o_{f_i} - \sum_R \nu_i \, \Delta H^o_{f_i} \quad ,$$

where

ν_i = the stoichiometric coefficient of species i

$H^o_{f_i}$ = the standard heat of formation (1 atm, 298 K) of species i

P = reaction products

R = reactants

$\Delta H^o_{f_{SO_2}}$ = -296.06 kJ/g · mol*

$\Delta H^o_{f_{O_2}}$ = 0.0*

$\Delta H^o_{f_{SO_3}}$ = -395.18 kJ/g · mole.*

Thus, ΔH^o_R = -395.18 - [-296.06] = -99.12 kJ/g · mol.

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*Barrow, G. M., Physical Chemistry, 3rd Ed. New York. McGraw Hill, 1973.

P.2 DETERMINATION OF THE HEAT OF REACTION AT THE TEMPERATURE OF THE CATALYST BEDS

Optimum conversion temperature ~438° C = 711 K

ΔH_{711} = heat of reaction at 711 K

$$\Delta H_{711} = \Delta H^{o}_{R} + \int_{298\ K}^{711\ K} \Delta C_{p}\, dT ,$$

where

$$\Delta C_{p} = \sum_{P} \nu_{i}\, C_{p_{i}} - \sum_{R} \nu_{i}\, C_{p_{i}}$$

$C_{p_{i}}$ = the specific heat capacity of species i

ν_{i} = the stoichiometric coefficient of species i

$C_{p_{SO_3}}$ = 50.63 J/g · mol °K*

$C_{p_{SO_2}}$ = 39.79 J/g · mol °K*

$C_{p_{O_2}}$ = 29.36 J/g · mol °K.*

Thus, $\Delta C_{p} = 50.63 - [(0.5)(29.36) + 39.79] = -3.84$ J/g · mol °K

$$\therefore \int_{298\ K}^{711\ K} \Delta C_{p} dT = -3.84\ [711 - 298] = -1{,}586 \text{ J/g} \cdot \text{mol} .$$

Thus,

$$\Delta H_{711} = -99.12 \frac{\text{kJ}}{\text{g} \cdot \text{mol}} \times \frac{10^{3}\ \text{J}}{\text{kJ}} - 1{,}586 \frac{\text{J}}{\text{g} \cdot \text{mol}}$$

$$= -100{,}706 \text{ J/g} \cdot \text{mol} .$$

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*Letter and attachments from Arzabe, H. A., Monsanto Enviro-Chem, to Wood, J. P., Research Triangle Institute. August 3, 1982. Response to request to review Draft Chapter 4 of BID and acid plant preheater operating cost estimation procedure.

P.3 CALCULATION OF THE HEAT DEFICIENCY THAT RESULTS WHEN THE GAS STREAM SO_2 CONCENTRATION FALLS BELOW THE AUTOTHERMAL REQUIREMENT

$$J = 251{,}140 \sum_{i=1}^{n} T_i V_i (C_A - C_i)$$

where,

J = the heat deficiency during a 24-hour cycle (kilojoules)

n = the number of time periods during a 24-hour cycle during which the gas stream SO_2 concentration is below that required for autothermal operation.

T_i = the duration of time period i (hours)

V_i = the gas stream volumetric flow rate in evidence during time period i (Nm^3/min)

C_A = the gas stream SO_2 concentration required to sustain autothermal operation (volume portion)

C_i = the gas stream SO_2 concentration in evidence during time period i (volume portion)

P.4 ILLUSTRATION OF THE COST ESTIMATION PROCEDURE

Consider the following 24-hour gas stream profile:

No. of hours per day	Gas stream volumetric flow rate (Nm^3/min)	C_i, %
4.0	6,690	3.8
5.2	5,860	3.2
9.5	5,830	3.6
2.3	5,000	2.8
1.7	4,980	3.2
1.3	4,146	2.3

A single contact/single absorption acid plant is specified. From the information presented above, it can be determined that $n = 4$. Thus,

$$1.2 \sum_{i=1}^{n} T_i V_i \, (C_A - C_i) \, \Delta H_{711}$$

$$= 1.2 \left[5.2 \text{ hrs} \times \frac{5{,}860 \text{ Nm}^3}{\text{min}} \times (0.035 - 0.032) + 2.3 \times 5{,}000 \right.$$

$$\times (0.035 - 0.028) + 1.7 \times 4{,}980 \times (0.035 - 0.032) + 1.3 \times 4{,}140$$

$$\left. \times (0.035 - 0.023) \right] \times \frac{60 \text{ min}}{\text{hr}} \times \frac{101 \text{ kJ}}{\text{g} \cdot \text{mol}} \times \frac{\text{g} \cdot \text{mol}}{0.02413 \text{ m}^3 \ (21^\circ \text{ C})} = 78.9$$

$$\times 10^6 \text{ kJ per 24 hours} .$$

Assuming that the heating value of natural gas is 37,228 kJ/m^3, that the cost of natural gas is \$97.82 per 10^3m^3, and that the facility operates at 8,400 hr/yr, the annual cost attributable to preheating requirements can be estimated as follows:

$$\frac{78.9 \times 10^6 \text{ kJ}}{24 \text{ hours}} \times \frac{\text{m}^3}{37{,}228 \text{ kJ}} \times \frac{\$97.82}{10^3\text{m}^3} \times \frac{8{,}400 \text{ hr}}{\text{yr}} = \$72{,}586 \text{ per year} .$$

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

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16. ABSTRACT

Standards of performance for the control of emissions from primary copper smelters were promulgated in 1976. Developments since promulgation necessitated that the following be included in the periodic review of the standards: (1) reexamination of the current exemption for reverberatory furnaces processing high-impurity materials, (2) assessment of the feasibility of controlling particulate matter emissions from reverberatory furnaces processing high-impurity materials, (3) reevaluation of the impact of the current standard on the ability of existing smelters to expand production, and (4) assessment of the technical and economic feasibility of controlling fugitive emissions at primary copper smelters. The results of the review indicated that no changes should be made to the existing standard. This document contains background information and environmental and economic assessments considered in arriving at this conclusion.

This report is published in two volumes. Volume 1, EPA 450/3-83-018a, contains Chapters 1 through 9. Volume 2, EPA 450/3-83-018b, contains the Appendixes.

17. KEY WORDS AND DOCUMENT ANALYSIS

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